U.S. DEPARTMENT OF COMMERCE
National Technical Information Service

PB-261 922

Desulfurization of Steel Mill Sinter Plant Gases

Radian Corp, Austin, Tex

Prepared for

Industrial Environmental Research Lab, Research Triangle Park, N C

Oct 76

012036

PB 261 922

EPA-600/2-76-281
October 1976

Environmental Protection Technology Series

DESULFURIZATION OF STEEL MILL SINTER PLANT GASES

REPRODUCED BY
NATIONAL TECHNICAL
INFORMATION SERVICE
U. S. DEPARTMENT OF COMMERCE
SPRINGFIELD, VA. 22161

Industrial Environmental Research Laboratory
Office of Research and Development
U.S. Environmental Protection Agency
Research Triangle Park, North Carolina 27711

PB-261 922

TECHNICAL REPORT DATA

(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA-600/2-76-281	2.	3. RECIPIENT'S ACCESSION NO.
4. TITLE AND SUBTITLE Desulfurization of Steel Mill Sinter Plant Gases		5. REPORT DATE October 1976 6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) Gary D. Brown, Richard T. Coleman, James C. Dickerman, and Philip S. Lowell		8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS Radian Corporation P.O. Box 9948 Austin, Texas 78766		10. PROGRAM ELEMENT NO. 1AB015; ROAP 21AQR-005 11. CONTRACT/GRANT NO. 68-02-1319, Task 58
12. SPONSORING AGENCY NAME AND ADDRESS EPA, Office of Research and Development Industrial Environmental Research Laboratory Research Triangle Park, NC 27711		13. TYPE OF REPORT AND PERIOD COVERED Task Final; 3-9/76 14. SPONSORING AGENCY CODE EPA-ORD

15. SUPPLEMENTARY NOTES IERL-RTP task officer for this report is Norman Plaks, 919/549-8411 Ext 2557, Mail Drop 62.

16. ABSTRACT The report gives results of an evaluation of the technical and economic feasibility of using limestone scrubbing technology to control sinter plant emissions. Data from Soviet and Japanese sinter plants employing limestone scrubbing technology were used to develop a realistic design basis. A conceptual process design was developed and used to prepare economic estimates. Results of the process design indicate that control of sinter plant emissions by limestone scrubbing is technically feasible. Economic evaluations show that limestone scrubbing will increase the cost of producing sinter by about $1.82 per metric ton of product sinter for a standard sinter plant operation. For a sinter plant with a windbox gas recirculation system, the cost increase would be about $1.44 per metric ton of product sinter.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group	
Air Pollution	Desulfurization	Air Pollution Control	13B	07A
Sintering Furnaces	Scrubbers	Stationary Sources	13A	
Iron and Steel Industry	Calcium Oxides	Windbox	11F	
Sulfur Oxides	Limestone	Gas Recirculation	07B	08G
Dust		Particulate	11G	
Cost Analysis		Lime/Limestone Scrubbing	14A	

18. DISTRIBUTION STATEMENT Unlimited	19. SECURITY CLASS *(This Report)* Unclassified	21. NO. OF PAGES 215
	20. SECURITY CLASS *(This page)* Unclassified	22. PRICE $7.75

EPA Form 2220-1 (9-73)

RESEARCH REPORTING SERIES

Research reports of the Office of Research and Development, U.S. Environmental Protection Agency, have been grouped into five series. These five broad categories were established to facilitate further development and application of environmental technology. Elimination of traditional grouping was consciously planned to foster technology transfer and a maximum interface in related fields. The five series are:

1. Environmental Health Effects Research
2. Environmental Protection Technology
3. Ecological Research
4. Environmental Monitoring
5. Socioeconomic Environmental Studies

This report has been assigned to the ENVIRONMENTAL PROTECTION TECHNOLOGY series. This series describes research performed to develop and demonstrate instrumentation, equipment and methodology to repair or prevent environmental degradation from point and non-point sources of pollution. This work provides the new or improved technology required for the control and treatment of pollution sources to meet environmental quality standards.

EPA REVIEW NOTICE

This report has been reviewed by the U.S. Environmental Protection Agency, and approved for publication. Approval does not signify that the contents necessarily reflect the views and policies of the Agency, nor does mention of trade names or commercial products constitute endorsement or recommendation for use.

This document is available to the public through the National Technical Information Service, Springfield, Virginia 22161.

EPA-600/2-76-281
October 1976

DESULFURIZATION OF STEEL MILL SINTER PLANT GASES

by

Gary D. Brown, Richard T. Coleman,
James C. Dickerman, and Philip S. Lowell

Radian Corporation
P.O. Box 9948
Austin, Texas 78766

Contract No. 68-02-1319, Task 58
ROAP No. 21AQR-005
Program Element No. 1AB015

EPA Task Officer: Norman Plaks

Industrial Environmental Research Laboratory
Office of Energy, Minerals, and Industry
Research Triangle Park, NC 27711

Prepared for

U.S. ENVIRONMENTAL PROTECTION AGENCY
Office of Research and Development
Washington, DC 20460

ABSTRACT

This report presents the results of a study to evaluate the technical and economic feasibility of using limestone scrubbing technology to control sinter plant emissions. Data from Soviet and Japanese sinter plants employing limestone scrubbing technology were used to develop a realistic design basis. A conceptual process design was developed and used to prepare economic estimates.

Results of the process design indicate that control of sinter plant emissions by limestone scrubbing is technically feasible. Economic evaluations show that a retrofitted limestone scrubbing system will increase the cost of producing sinter by about \$2.07 per metric ton of product sinter for a standard sinter plant operation. For a sinter plant with a windbox gas recirculation system the cost increase would be about \$1.59 per metric ton of product sinter.

TABLE OF CONTENTS

		Page
	ABSTRACT	iii
	TABLE OF CONTENTS	iv
	LIST OF TABLES	vi
	LIST OF FIGURES	ix
	CONVERSION FACTORS	x
	GLOSSARY	xi
1.0	SUMMARY	1
2.0	INTRODUCTION	3
3.0	TECHNICAL DISCUSSION	5
3.1	Description of Steel Mill Sinter Plants	5
3.1.1	Process Description	6
3.1.2	Process Developments	9
3.1.3	Sinter Plant Emissions	13
3.2	Description of the Lime/Limestone Wet Scrubbing Process	18
3.2.1	Process Description	20
3.2.2	Design Considerations	25
3.2.3	Typical Operations Relating to Sinter Plants	32
3.3	Evaluation of USSR and Japanese Data	33
3.3.1	Summary of Soviet Data	33
3.3.2	Summary of Japanese Data	35

TABLE OF CONTENTS (cont.) Page

4.0 DESIGN APPROACH.......... 37
4.1 Design Basis and Assumptions.......... 38
4.1.1 Sinter Plant Design Basis.......... 39
4.1.2 Limestone Scrubbing Design Basis.......... 44
4.2 Economic Basis.......... 55
4.2.1 Capital Investment Costs.......... 55
4.2.2 Annual Operating Costs.......... 58

5.0 RESULTS.......... 60
5.1 Process Designs.......... 60
5.2 Limestone Scrubbing System Layout.......... 69
5.3 Economic Evaluation.......... 74
5.3.1 Total Capital Investment.......... 76
5.3.2 Annual Operating Costs.......... 78

6.0 CONCLUSIONS AND RECOMMENDATIONS.......... 81
6.1 Conclusions.......... 81
6.2 Recommendations.......... 82

7.0 REFERENCES.......... 85

APPENDIX A - COMMENTS ON THE SOVIET DATA.......... A-1
APPENDIX B - COMMENTS ON THE JAPANESE DATA.......... B-1

TABLE OF CONTENTS (cont.) Page

APPENDIX C - DESCRIPTION OF RADIAN'S PROCESS SIMULATION MODEL.......... C-1

APPENDIX D - PROCESS EQUIPMENT LIST AND COST DATA.............................. D-1

LIST OF TABLES

Page

TABLE 3-1 SINTER-MIX COMPOSITION........................ 8

TABLE 3-2 MATERIALS USED IN THE PRODUCTION OF SINTER AT STEEL PLANTS IN THE UNITED STATES......... 12

TABLE 3-3 COMPOSITION OF PARTICULATE EMISSIONS......... 15

TABLE 3-4 SIZE DISTRIBUTION OF PARTICULATE EMISSIONS... 16

TABLE 3-5 TYPICAL CONCENTRATIONS OF GASEOUS EMISSIONS FROM STEEL MILL SINTER PLANTS................ 17

TABLE 3-6 SULFUR BALANCE FOR SINTER MACHINE OPERATION.. 19

TABLE 3-7 COMPARISON OF SCRUBBER TYPES FOR A LIMESTONE WET SCRUBBING SYSTEM........................ 28

TABLE 4-1 SINTER PLANT DESIGN BASIS.................... 40

TABLE 4-2 COMPOSITION OF PARTICULATES IN SINTER PLANT FLUE GAS.................................... 43

TABLE 4-3 LIMESTONE COMPOSITION........................ 45

TABLE 4-4 MAKEUP WATER COMPOSITION..................... 45

TABLE 4-5 LIMESTONE SCRUBBING DESIGN PARAMETERS........ 50

TABLE 4-6 MASS EMISSION RATES FROM THE RADIAN BASE CASE STEEL MILL SINTER PLANT AFTER LIMESTONE SCRUBBING OF THE WINDBOX EXHAUST GAS.............. 52

TABLE 4-7 ITEMS USED TO ESTIMATE THE TOTAL CAPITAL INVESTMENT REQUIRED FOR A LIMESTONE SLURRY PROCESS.................................... 57

TABLE 4-8 BREAKDOWN OF ANNUAL OPERATING COSTS FOR A LIMESTONE SLURRY PROCESS..................... 59

TABLE 5-1 MATERIAL BALANCE FOR A LIMESTONE SCRUBBING PROCESS ON A STANDARD SINTER PLANT........... 63

LIST OF TABLES (cont.) Page

TABLE 5-2 MATERIAL BALANCE FOR A LIMESTONE SCRUBBING PROCESS ON A SINTER PLANT WITH WINDBOX RECYCLE.................................... 64

TABLE 5-3 DESCRIPTION AND DESIGN SPECIFICATIONS FOR MAJOR PROCESS EQUIPMENT..................... 65

TABLE 5-4 OPERATING PARAMETERS FOR PROCESS DESIGNS..... 68

TABLE 5-5 SPACE REQUIREMENTS FOR A LIMESTONE SCRUBBING SYSTEM ON STEEL MILL SINTER PLANT APPLICATIONS 73

TABLE 5-6 TOTAL CAPITAL INVESTMENT SUMMARY FOR STEEL MILL SINTER PLANT FLUE GAS DESULFURIZATION USING LIMESTONE SLURRY SCRUBBING............ 77

TABLE 5-7 LIMESTONE SLURRY PROCESS TOTAL ANNUAL OPERATING COSTS (STANDARD CASE)................... 79

TABLE 5-8 LIMESTONE SLURRY PROCESS TOTAL ANNUAL OPERATING COSTS (RECYCLE CASE).................... 80

LIST OF FIGURES

Page

FIGURE 3-1 SCHEMATIC FLOW DIAGRAM FOR TYPICAL MODERN SINTER PLANT.................................. 10

FIGURE 3-2 PROCESS FLOW DIAGRAM - LIME/LIMESTONE WET SCRUBBING 21

FIGURE 5-1 PROCESS FLOW DIAGRAM - LIMESTONE SCRUBBING PROCESS FOR STEEL MILL SINTER PLANT APPLICATION................................ 62

FIGURE 5-2 LAYOUT OF SCRUBBING SECTION OF A LIMESTONE SCRUBBING PROCESS FOR A STANDARD STEEL MILL SINTER PLANT OPERATION...................... 70

FIGURE 5-3 LAYOUT OF SCRUBBING SECTION OF A LIMESTONE SCRUBBING PROCESS FOR A RECYCLE STEEL MILL SINTER PLANT OPERATION...................... 71

FIGURE 5-4 LAYOUT OF FEED PREPARATION AND SLURRY PROCESSING SECTION OF A LIMESTONE SCRUBBING PROCESS FOR A STEEL MILL SINTER PLANT APPLICATION................................ 72

CONVERSION FACTORS

The metric system is used in this report. Following are some factors for conversion between metric and English systems:

1 m (meter) = 3.281 feet
1 m^3 (cubic meter) = 35.314 cubic feet
1 mt (metric ton) = 1.1023 short tons
1 kg (kilogram) = 2.2046 pounds
1 liter = 0.2642 gallon

The capacity of FGD systems is expressed in Nm^3/hr (normal cubic meters per hour)

1 Nm^3/hr = 0.589 SCFM

L/G ratio (liquid/gas ratio) is expressed in liters/Nm^3

1 liter/Nm^3 = 7.481 gallons/1000 SCF

GLOSSARY

Carbide Sludge - Calcium oxide (CaO) and impurities formed as a byproduct of acetylene manufacture.

Coke Breeze - coke fines that are generated during the crushing and sizing of the coke for blast furnace consumption and which are used as a fuel in the sinter charge.

Flue Gas - sinter plant off-gas.

Fluxing - any process in which materials (Fluxes) are added to the metal charge to aid in the removal of gases, oxides, or other impurities.

Fluxstone - limestone or dolomite.

Fly Ash - particulates entrained in the sinter plant off-gas.

Gangue - a waste rock or slag material remaining after most of the metal values have been removed.

Relative Saturation - The relative saturation (RS) is the product of the activities of the species which react to produce the precipitating solid divided by the solubility product constant, as shown in the following equations for calcium sulfate.

$$Ca^{++} + SO_4^{=} + 2H_2O \rightleftharpoons CaSO_4 \cdot 2H_2O \downarrow$$

$$RS = \left[a_{Ca^{++}} \cdot a_{SO_4^{=}} \cdot a^2_{H_2O(\ell)}\right] / K_{sp,CaSO_4 \cdot 2H_2O}$$

For precipitation to occur, the relative saturation must be greater than one, and the rate R positive (see equation 2-4, Appendix C). For dissolution to occur, the relative saturation must be less than one, and the rate R negative. At equilibrium, the relative saturation is equal to one, and the rate is zero.

Slag - A residue that forms on the surface of molten metal during fluxing.

1.0 SUMMARY

Desulfurization of sinter plant windbox gases by limestone scrubbing is commercially practiced in both the USSR and Japan; it is not currently practiced in the United States. To determine the costs of applying this technology domestically, conceptual process designs were prepared for both standard sinter plant operations and operations employing a windbox gas recirculation system. Results of the conceptual designs were used to size process equipment. An economic basis was selected and applied to the process designs to perform an evaluation of both capital and operating costs for each system.

The services of two outside consultants were retained in order to help in obtaining the data and information that was necessary to perform this evaluation. Mr. Richard Jablin, a consultant with over 35 years experience in steel mill engineering and environmental control, provided much assistance and information on steel mill sinter plant operations. Mr. Jablin holds several patents in the area of steel making and has been employed by various steel companies since 1950. He currently directs a consulting engineering firm, Richard Jablin and Associates, located in Winchester, Virginia. Dr. Jumpei Ando, an international consultant and lecturer in the areas of SO_x and NO_x control, provided data and descriptions of several lime and limestone systems which are currently being used in Japan to remove SO_2 from sinter plant gases. Dr. Ando has held previous positions with the Faculty of Engineering, University of Tokyo and the Tennessee Valley Authority. He is currently a Professor at Chuo University in the Faculty of Science and Engineering.

The process designs and cost estimates are based upon data obtained from the following sources:

1) information available from Radian files, Mr. Richard Jablin, and the open literature;

2) information on Soviet technology obtained by the EPA as a result of a technology interchange agreement between the US and USSR;

3) information on Japanese technology prepared for Radian by Dr. Jumpei Ando;

4) cost data provided in a TVA report prepared by McGlamery, et al. (MC-147).

The results of this evaluation indicate that the capital costs of a limestone scrubbing system, applied to a sinter plant having a capacity of 6312 mtpd of product, range from $8-10 million, depending upon whether or not the sinter plant uses windbox gas recirculation. For the same sinter plants the operating costs would be respectively, $1.59-2.07 per metric ton of product sinter. The desulfurization system evaluated here uses a venturi prescrubber which would effectively remove particulates. Optimization of the technology could be accomplished by determining the effects of contaminants in the sinter plant windbox gases on limestone scrubbing process chemistry and on corrosion in the prescrubber loop.

2.0 INTRODUCTION

Sulfur dioxide (SO_2) emissions from steel mill sinter plants are of concern to the U.S. Environmental Protection Agency. Past experience of both the U.S. electric power industry and of USSR and Japanese steel manufacturers indicates that limestone slurry scrubbing is a feasible technique for sinter plant emission control. Under contract to EPA, Radian has completed this study to evaluate the applicability of limestone slurry scrubbing to the sinter plant emission control problem.

The objectives of the study were twofold. First, the evaluation was performed to determine the technical feasibility of applying limestone wet scrubbing technology to control sinter plant emissions. Potential process problems and land requirements for solid waste disposal were identified. Secondly, the study was to provide EPA with process economics to aid them in determining the economic feasibility of applying limestone technology to sinter plants.

To accomplish these objectives, the following approach was taken. First, data on U.S. sinter plant operations were collected and reviewed. Secondly, data from USSR and Japanese sources were collected and evaluated to determine the applicability of foreign experience to domestic applications. From these evaluations, a design basis was formulated and used to develop conceptual process designs for limestone scrubbing systems to control sinter plant emissions. Both standard sinter plant operations and windbox gas recirculation operations were used as design cases. Finally, estimates were made of process economics for both design cases.

The following sections of this report describe the evaluations performed as part of this study. Section 3.0 is a technical discussion of sinter plant and limestone scrubbing operations. A summary of the evaluations of Soviet and Japanese experiences with limestone scrubbing of sinter plant emissions is also included. Section 4.0 provides details of the design approach and basis. Section 5.0 reports the study results and Section 6.0 presents conclusions and recommendations. Supporting details are included in the appendices.

3.0 TECHNICAL DISCUSSION

In order to evaluate the applicability of limestone slurry scrubbing to steel mill sinter plants, information on both the operations and emissions of U.S. sinter plants was collected and reviewed. Information on limestone slurry scrubbing operations, obtained from previous Radian reports and in-house files, was used as a basis for determining the process design considerations that would be important as related to sinter plant applications. Data were also collected from Japanese and Russian sources on the operating characteristics of limestone slurry scrubbing units which currently process steel mill sinter plant flue gases.

Typical sinter plant operations and emissions are described in Section 3.1. A description of limestone slurry scrubbing is given in Section 3.2 and a summary of the data collected from both Soviet and Japanese limestone scrubbing experiences is presented in Section 3.3.

3.1 Description of Steel Mill Sinter Plants

The following sinter plant description has been taken largely from a report prepared for EPA by National Steel Corporation entitled Sinter Plant Windbox Gas Recirculation System Demonstration (PE-179).

The function of sintering, in the steel industry, is to convert iron-bearing raw materials of a fine particle size into coarse agglomerates by partial fusion. The sinter product has a porous cellular structure resembling clinker in physical appearance. Its composition may be substantially different from that of the original iron-bearing fines.

Blast furnace sinter is categorized as acidic or basic, depending on the basicity ratio. The basicity ratio is defined by the following equation:

$$\text{Basicity Ratio} = \frac{\text{Wt Percent CaO} + \text{Wt Percent MgO}}{\text{Wt Percent SiO}_2 + \text{Wt Percent Al}_2\text{O}_3}$$

Sinter with a basicity ratio of less than 1.0 is acid and that with a ratio greater than 1.0 is basic. It has become common to refer to sinter with a basicity ratio of approximately 1.0 as self-fluxing, while ratios in excess of 1.0 may be called burden-fluxing or superfluxed sinter. Acid sinter of basicity less than 0.5 was the predominant product used as blast furnace feed until the early 1950's. It was then realized that both economic and productive benefits to the blast furnace could be realized by incorporating in the sinter a part or all of the required furnace flux. This was achieved by the addition of limestone and/or dolomitic fines to the ore fines to be sintered.

3.1.1 Process Description

In the sintering process, a shallow bed of fine particles is agglomerated by heat exchange and partial fusion of the quiescent mass. Heat is generated by the combustion of a solid fuel contained within the bed of fines being agglomerated. The process is initiated by igniting the fuel at the top surface of the bed, after which a thin, high temperature combustion zone is drawn downward through the bed by an induced draft. Within this zone, the surfaces of adjacent particles are at a fusion temperature and the gangue constituents form a semiliquid slag. The flow of volatiles and incoming air creates a frothy condition and freezes the trailing edge of the advancing fusion zone. The product then consists of particles of ore bonded in a slag matrix of cellular structure.

In the ferrous industry, the material to be sintered consists essentially of a mixture of iron-bearing fines and a solid fuel. The iron-bearing constituents are chiefly iron ore fines, recycled sinter fines and blast furnace flue dust, but may also include mill scale, and other steel mill waste products containing iron. Coke breeze is the most common solid fuel, but other carbonaceous materials are used. It has become common practice to add limestone or dolomitic fines to the sinter mix and this material may now be considered as an essential constituent in a typical sinter mix. Sinter-mix compositions for three different U.S. steel mill sinter plants are shown in Table 3-1 (VA-126). This mixture of fine material is placed on the sinter strand in a shallow bed, seldom less than 152 mm (6 inches) nor more than 508 mm (20 inches) in depth. In the ignition zone, the surface of the bed is heated to about 1260 to 1371°C (2300-2500°F), combustion of the fuel is initiated, and fusion of the fine particles at the surface begins. As air is drawn through the bed, the high temperature zone of combustion and fusion moves downward through the bed and produces the bonded, cellular structure.

Combustion of the solid fuel and propagation of the fusion zone through the bed is dependent on the air flow. To assure an adequate air flow, the sinter mix is generally preconditioned to improve its permeability. This can be accomplished by eliminating excessively fine materials when economically possible, but is normally achieved by the addition of coarse [12.7 mm by 3.2 mm (½ inch by ⅛ inch)] sinter returns and by fluffing and/or micropelletizing the fine particles in a balling drum.

The design and physical arrangement of sintering equipment and the flow pattern of raw materials and product

TABLE 3-1

SINTER-MIX COMPOSITION

(weight percent)

Sinter-Mix Components	Plant 1	Plant 2	Plant 3
Iron Ore	29.1	82.0	49.5
Dry blast-furnance dust	1.3		5.6
Blast-furnance filter cake	15.8		4.9
Melt-shop slag	6.5		0.0
Rolling-mill scale	11.5		7.6
Basic-oxygen-furnance dust	3.5	0.0	0.0
Miscellaneous dust	0.0	0.0	6.7
Limestone or dolomite	28.2	15.0	20.8
Coke	4.1	3.0	4.9
Total	100.0	100.0	100.0

Source: VA-126

differ considerably between various plants. The choice of equipment and material flow is generally determined by the desired capacity, space available, capital costs, and prevailing technology. However, each plant can be divided into three distinct phases of operation, namely, (1) raw materials processing, (2) sinter production and (3) product processing. A schematic flow diagram of a typical modern sinter plant is shown in Figure 3-1.

This schematic flow diagram is typical of a modern sinter plant. Many of the older plants lack certain features, such as hot and/or cold screens, balling facilities, sinter breakers, coolers, and flexibility in materials handling. Some of the newest plants have special process control features to reduce the variation in the sinter product. Examples of the new control technology being applied to sinter plants are: (1) automatic raw material proportioning systems, and (2) computer control of the process.

A typical sinter plant strand normally operates at full load except for start-up. The induced draft fan that pulls air through the sinter bed maintains a uniform suction pressure. This results in a fairly uniform gas flow through the sinter bed. The frequency of shutdown depends on the condition of the plant. In a well run sinter plant the operator would plan on running all week without an unscheduled shutdown. Scheduled shutdowns for a well run sinter plant are about one eight-hour shift every week (JA-136).

3.1.2 Process Developments

Most sinter plants are operated in a similar manner, differing primarily in the characteristics of the raw materials which must be processed and the basicity ratio chosen for the sinter product. The general trend in materials used for sin-

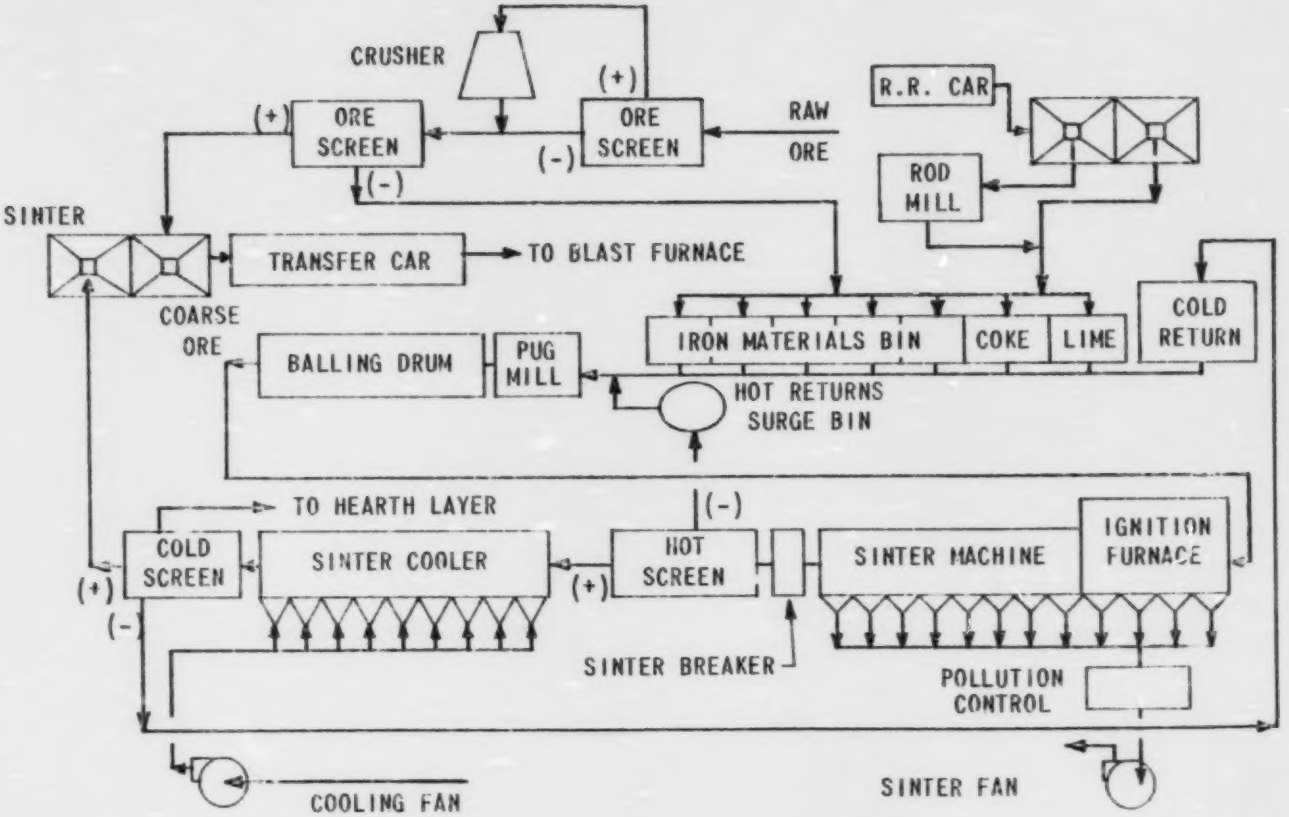

CRUSHER
(+)
ORE SCREEN
(+)
ORE SCREEN
(-)
RAW ORE
R.R. CAR
(-)
ROD MILL
SINTER
TRANSFER CAR
TO BLAST FURNACE
COARSE ORE
IRON MATERIALS BIN
COKE
LIME
COLD RETURN
BALLING DRUM
PUG MILL
HOT RETURNS SURGE BIN
TO HEARTH LAYER
(-)
COLD SCREEN
SINTER COOLER
HOT SCREEN
SINTER MACHINE
IGNITION FURNACE
(+)
(+)
(-)
SINTER BREAKER
POLLUTION CONTROL
COOLING FAN
SINTER FAN

Figure 3-1 Schematic flow diagram for typical modern sinter plant

tering has been toward less iron ore fines and more iron-bearing waste materials such as mill scale, ironmaking dusts and slags. Fluxstone additions have increased with the trend toward higher sinter basicities. Table 3-2 shows the trend in sinter feed materials between 1960 and 1968. The increased use of mill scale, cinder and slag, other materials which are primarily recyclable iron-bearing fines, and fluxstone is evident. The reduction in the use of blast furnace flue dust and sludge is due to improved blast furnace burdens.

A major factor influencing sinter plant operation has been the trend toward higher sinter basicities. In 1962, only about 40 percent of the sinter produced had a basicity ratio in excess of 1.0. At the present time, available data indicate that at least 85 percent of the sinter produced in the United States and Canada has a basicity in excess of 1.0. Moreover, at least six sinter plants regularly produce sinter with a basicity ratio greater than 3.0 and one plant in excess of 4.0. Increasing the sinter basicity generally reduces production capacity but increases the strength of the sinter product. Improved sinter quality reduces the quantity of fines to be recycled in the sinter plant and decreases the flue dust and sludge generated at the blast furnace. Higher basicity sometimes improves sinter strand operation and reduces the production and emission of fine dusts.

A new development in sinter plant operations is the practice of recirculating a portion of the windbox flue gas. Development work, including installation and operation of a windbox recirculation system, has been performed by the National Steel Corporation, Wierton Steel Division, and the Aliquippa Works of Jones and Laughlin Steel Corporation. The recirculating system at Wierton is designed with four duct valves which permit a once-through operation of recirculation rates ranging from 0-50 percent. A recirculation rate of 39 percent was calculated to be the maximum recycle amount because of the rapidly decreasing oxygen content in the recirculated gas, and the large

TABLE 3-2

MATERIALS USED IN THE PRODUCTION OF SINTER AT STEEL PLANTS IN THE UNITED STATES

		Year		
MATERIAL		1960	1964	1968
Iron Ore	a)	35,900	39,370	33,170
	b)	36,500	40,000	33,700
	c)	74	69	64
Flue Dust & Sludge		5,200	4,530	3,050
		5,300	4,600	3,100
		11	8	6
Scale		980	2,170	3,150
		1,000	2,200	3,200
		2	4	6
Cinder and Slag		49	394	394
		50	400	400
		.01	1	1
Other		490	1,080	1,770
		500	1,100	1,800
		1	2	3
Fluxstone		3,740	5,800	7,480
		3,800	5,900	7,600
		7	10	15
Fuel		2,360	3,250	2,660
		2,400	3,300	2,700
		5	6	5

a) Thousands of metric tons

b) Thousands of gross tons

c) Percentage of mix

Source: PE-179

rate of increase in fan horsepower with increasing recycle percentage.

The windbox recirculation system offers the following potential advantages as reported by Wierton (PE-179):

(1) reduction in quantity of gases to be cleaned,

(2) reduction in capital investment for gas cleaning equipment,

(3) reduction in total emissions to the atmosphere for a given dust concentration,

(4) reduction in hydrocarbon content of the exhaust gases due to more complete combustion, and

(5) conservation of energy from recirculating hot gases.

Because of the potential advantages to be gained from operating in a windbox recirculation mode, data from the Wierton Steel Division operations have been used to prepare a conceptual design of a limestone scrubbing system to treat SO_2 emissions. Results of this conceptual design will be compared against a design of a limestone system for treating a sinter plant waste stream without windbox recirculation.

3.1.3 Sinter Plant Emissions

Before discussing sinter plant emissions it should be mentioned that every sinter plant is a special case. They are all different, having different feed compositions, making it difficult to define a typical sinter plant.

Emissions from sinter plants include particulates, condensable aliphatic hydrocarbons, and gaseous components such as sulfur dioxide, carbon monoxide, and chlorides. Table 3-3 shows the particulate compositions for the three steel mill sinter plants whose sinter mix compositions were given in Table 3-1. The size distribution for particulates emitted from the main exhaust system of several sinter plants is given in Table 3-4.

Gaseous emissions from sinter plants include significant amounts of carbon monoxide and sulfur oxides. Typical concentrations of gaseous emissions from steel mill sinter plants are given in Table 3-5. Other gaseous components that are present in smaller amounts include nitrogen oxides, chlorides, and fluorides.

Emissions from a sinter plant with windbox recirculation vary from those of a sinter plant without recirculation. The major environmental effects of gas recirculation are an increase in the concentration of SO_2 in the exhaust gas, and a decrease in both hydrocarbon and particulate emissions. In both modes of sinter plant operation, the same quantity of sulfur is oxidized to produce SO_2. Therefore, although the concentration of SO_2 will increase in the recirculation case, the total quantity of SO_2 emitted for both cases will remain the same. The advantage gained by using a gas recirculation system is that the total gas volume to be processed is reduced thereby reducing the size and capital and operating cost of required emission control equipment.

Hydrocarbon emissions will be reduced as the hydrocarbons in the recycle stream will pass back through the flame zone to be combusted. The particulate concentration should be the same for both cases, although the windbox recirculation case will produce less total particulate emissions. The stream compositions used as a basis for the limestone scrubbing design are presented and discussed in Section 4.1 - Design Basis and Assumptions.

TABLE 3-3
COMPOSITION OF PARTICULATE EMISSIONS
(weight percent)

Particulate Component	Plant 1	Plant 2	Plant 3
Fe_2O_3	33.9	11.7	28.0
CaO	7.1	10.9	15.0
MgO	5.3	0.4	2.0
K_2O	5.2	0.6	8.1
SiO_2	4.8	2.4	4.6
Al_2O_3	2.6	4.3	2.5
Na_2O	1.6	0.3	0.0
ZnO	0.4	0.1	0.0
MnO	0.2	0.1	0.0
Chlorides	8.5	3.0	8.8
Sulfates	7.5	16.5	2.1
Hydrocarbons	7.4	36.9	0.0
Other	1.6	0.0	0.0
Loss on Ignition	13.9	12.3	28.9
Total	100.0	100.0	100.0

Source: VA-126

TABLE 3-4

SIZE DISTRIBUTION OF PARTICULATE EMISSIONS

Size grading of dust from sinter plant main exhaust gases, % undersize

	Size, microns										
Plant	295	211	152	104	76	53	40	30	20	10	5
A				89			50	41	33	19	7.5
				63	51	43	39	34	26	14	5
				78	65	55	50	44	33	17	7
B	84	71	54	37	25	18	16	14	10	6	2
C	95	90	79	58	33	18	14	8	6	2	--
	96	88	73	52	38	31	25	21	13	5	--
D	75	64	53	40	29	22	17	13	8	3	--
H			96	88	76	61	46	35	25	13	6

Source: BA-449

TABLE 3-5
TYPICAL CONCENTRATIONS OF GASEOUS EMISSIONS FROM STEEL MILL SINTER PLANTS

Component	Mole Percent
N_2	72.4
O_2	14.5
CO	0.7
CO_2	6.3
H_2O	6.1
SO_2	25 - 1000*
Condensable Hydrocarbons	693**

* ppm

** mg/Nm^3

Source: PE-179

A sulfur balance for a sinter-machine operation is given in Table 3-6. The major sources of sulfur are the iron-bearing materials and the coke. The fuel oil used for igniting the sinter mix and the limestone used as a flux also contain some sulfur. Sulfur is carried out of the system with the product sinter and as SO_2 in the combustion gases. It was estimated that aobut 36 percent of the sulfur entering with the sinter feed left in the combustion gases (VA-003). It should be realized that the sulfur balance presented here was made for a specific case. Sinter feeds with different compositions, basicity ratio, and using a different type of fuel for igniting the sinter bed can have a sulfur balance that is much different.

3.2 Description of the Lime/Limestone Wet Scrubbing Process

The lime/limestone flue gas desulfurization process uses a slurry of calcium oxide or calcium carbonate to absorb SO_2 in a wet scrubber. This process is commonly referred to as a "throwaway" process because the calcium sulfite and sulfate formed in the system are disposed of as waste solids. The overall reactions in the system are as follows.

For lime systems:

$$SO_{2\,(g)} + CaO_{(s)} + \tfrac{1}{2}H_2O \rightarrow CaSO_3 \cdot \tfrac{1}{2}H_2O_{(s)} \qquad (3\text{-}1)$$

For limestone systems:

$$SO_{2\,(g)} + CaCO_{3\,(s)} + \tfrac{1}{2}H_2O \rightarrow CaSO_3 \cdot \tfrac{1}{2}H_2O_{(s)} + CO_{2\,(g)} \qquad (3\text{-}2)$$

TABLE 3-6

SULFUR BALANCE FOR SINTER MACHINE OPERATION

Basis: The production of one metric ton of sinter

Material	Amount (kg)	Sulfur Content (wt. percent)	Amount of Sulfur (kg)
INPUT:			
Iron-bearing Material	1,100	0.041	0.45
Coke	50	0.70	0.35
Oil	25	0.55	0.14
Limestone	100	0.049	0.05
			0.99
OUTPUT:			
Sinter	1,000	0.055	0.55
Sinter Fines	144.5	0.055	0.08
Sulfur in Combustion Gases			0.36
			0.99

Source: VA-003

Some oxygen will also be absorbed from the flue gas or surrounding atmosphere and will cause oxidation of absorbed SO_2 and formation of calcium sulfate.

For limestone systems:

$$Ca^{++} + SO_3^{=} + \frac{1}{2}O_2 + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O_{(s)} \qquad (3\text{-}3)$$

The calcium sulfite and sulfate crystals are precipitated in a hold tank and then sent to a solid/liquid separator where the solids are removed. The waste solids are generally disposed of by ponding or landfill.

3.2.1 Process Description

The basic design of a lime or limestone scrubbing system can be divided into the following process areas:

(1) SO_2 Absorption,
(2) solid separation, and
(3) solids disposal.

Figure 3-2 shows a generalized process flow diagram for the lime/limestone slurry scrubbing processes.

SO_2 Absorption

Absorption of SO_2 takes place in a wet scrubber using lime or limestone in a circulating slurry. Carbide sludge (impure slaked lime) has also been used successfully at two installations. Particulates can be removed in the SO_2 absorber or ahead of the absorber by an electrostatic precipitator or particulate scrubber. The selection of a method for removal of particulates is based on economics and operational reliability. Removing particulates in the SO_2 absorber increases the solids

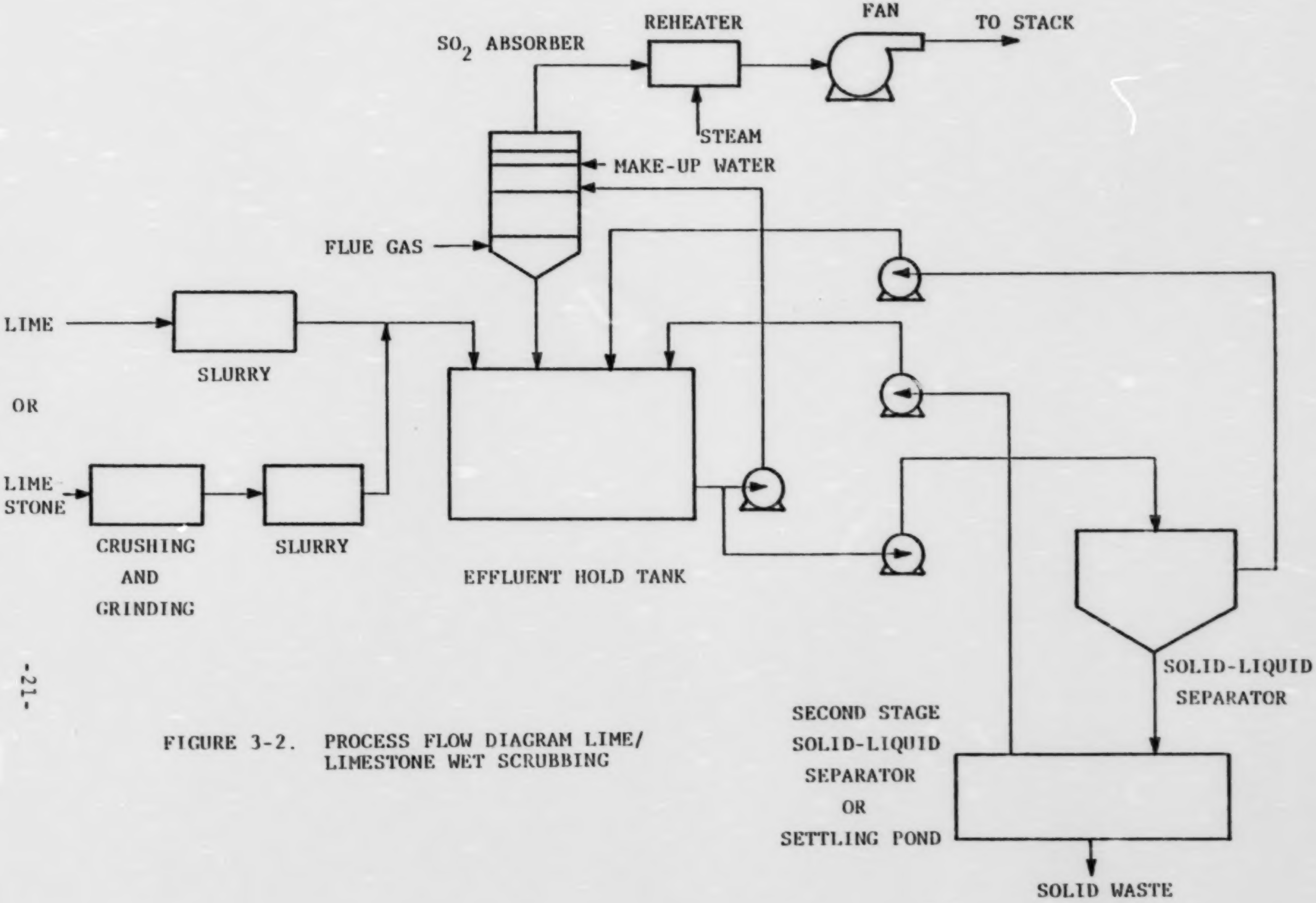

SO$_2$ ABSORBER
REHEATER
FAN
TO STACK
STEAM
MAKE-UP WATER
FLUE GAS
LIME
SLURRY
OR
LIME STONE
CRUSHING AND GRINDING
SLURRY
EFFLUENT HOLD TANK
SOLID-LIQUID SEPARATOR
SECOND STAGE SOLID-LIQUID SEPARATOR OR SETTLING POND
SOLID WASTE

FIGURE 3-2. PROCESS FLOW DIAGRAM LIME/ LIMESTONE WET SCRUBBING

load in the SO_2 scrubbing system. It is also believed that some components of the fly ash catalyze the oxidation of sulfite to sulfate which increases the potential for sulfate scaling.

The absorption of SO_2 from the flue gases by a lime or limestone slurry constitutes a multiphase system involving gas, liquid, and several solids. The overall reaction of gaseous SO_2 with the alkaline slurry yielding calcium sulfite and sulfate has been shown in Equations 3-1, 3-2, and 3-3. The solid sulfite is only very slightly soluble in the scrubbing liquor and thus will precipitate to form an inert solid for disposal. In the lime system some CO_2 may also be absorbed from the flue gas and will react in a similar fashion to form solid calcium carbonate.

In most cases some oxygen will also be absorbed from the flue gas or surrounding atmosphere. This leads to oxidation of absorbed SO_2 and precipitation of solid calcium sulfate as was shown in reaction 3-3.

The extent of oxidation can vary considerably, normally ranging anywhere from almost zero to 40 percent in the electric utility industry. In some systems treating dilute SO_2 flue gas streams, sulfite oxidation rates as high as 90 percent have been observed. In sintering operations, where the oxygen content of the flue gas is as high as 16 volume percent, sulfite oxidation rates of 100 percent have been reported. The actual mechanism for sulfite oxidation is not completely understood. The rate appears to be a strong function of oxygen concentration in the flue gas and liquor pH. It may also be increased by trace quantities of catalysts in fly ash entering the system.

Various types of gas-liquid contactors can be used as the SO_2 absorber. These differ in SO_2 removal efficiency as well as operating reliability. Four general types of contactors are usually used for SO_2 removal:

(1) venturi scrubbers,
(2) spray towers (horizontal and vertical),
(3) grid towers, and
(4) mobile bed absorbers [such as marble bed and turbulent contact absorber (TCA)].

The liquid to gas ratio (L/G) generally ranges between 4.7-14.7 liters/Nm^3 (35-110 gal/1000 scf) depending upon the type of contactor. Simple impingement devices are placed downstream from the absorber to remove mist entrained in the flue gas.

The effluent hold tank receives the lime or limestone feed slurry and absorber effluent. In addition, settling pond water and clarifier overflow can be sent to the hold tank. The volume of the hold tank is sized to allow residence time for adequate calcium sulfite and sulfate precipitation. Reaction time outside the scrubber is needed to allow the supersaturation caused by SO_2 sorption in the sorber to dissipate and to permit dissolution of absorbent. Too little residence time in the hold tank can cause calcium sulfite or sulfate scaling in the system.

The feed material for a lime scrubbing process is usually produced by calcining limestone. Feed for a limestone process generally comes directly from the quarry, and is then reduced in size by crushing and grinding. The lime or limestone is mixed with water to make a 25-60 percent solids slurry.

Solids Separation

A continuous stream of slurry of 10-15 percent solids is recycled to the absorber from the effluent hold tank. In addition, a bleed stream is taken off to be dewatered. The dewatering step, which is needed to minimize the area needed for sludge disposal, varies depending on the application and type of disposal. The waste sludge contains some unutilized lime or limestone. This depends upon system design (additive) stoichiometry). Generally, more excess additive is required in limestone systems than in lime systems.

For systems with on-site pond disposal, solids may be pumped directly from the effluent hold tank to the pond area. Clean overflow liquor from the pond would then be returned to the system. Depending on the physical properties of the solids produced in the system, a thickening device such as a clarifier can be used to increase the solids content to a maximum of about 40 weight percent. Additional dewatering to 60-70 percent solids can sometimes be achieved by vacuum filtration.

Solids Disposal

Sludge disposal is one of the main disadvantages of lime/limestone FGD systems in comparison to regenerable FGD processes. The quantity of sludge produced is large in both weight and volume, and requires a large waste pond or landfill area for disposal.

On-site disposal is usually performed by sending the waste solids to a large pond. Settling of the solids occurs and pond water is recycled back to the process hold tank for reuse. "Stabilization" methods are currently under development to convert the sludge to structurally-stable, leach-resistant, landfill material. These methods could be used when on-site disposal is not possible. The stabilized material can then be trucked to an off-site area for landfill.

At least four companies are developing sludge fixation processes. A fixation process is currently employed to dispose of the sludge generated by a limestone wet scrubbing system installed on a 163 Mw unit at Commonwealth Edison's Will County Station. The annual cost for sludge fixation is likely to be higher than the lime or limestone raw material cost. Conversion of the sludge to a construction material is another disposal method under consideration.

3.2.2 Design Considerations

The flow rate and sulfur content of the sinter plant flue gas are the major parameters to be considered in the lime or limestone scrubbing system design. The quantities of lime or limestone consumed and waste solids produced are roughly proportional to the amount of sulfur in the gas. The tendency for scale formation in the system is also related to the amount of SO_2 removed from the gas. Since all of the SO_2 removed must precipitate from solution before leaving the system, increased crystal seed must be provided in proportion to the amount of SO_2 removed. The scrubber liquid to gas ratio would have to be increased for removal of high SO_2 concentrations and to avoid exceeding the scaling limits in the scrubber effluent liquor.

Gas-Liquid Contactors

The different types of gas-liquid contactors can be separated into two categories; those having an open configuration and those having a closed (packed) configuration. These gas-liquid contactors differ in their gas velocity, L/G ratio, gas-side pressure drop, and resistance to plugging. These characteristics will be discussed for the different scrubber types.

Depending on the scrubber type, SO_2 removal efficiency may be increased by:

(1) increasing the number of scrubber stages,
(2) increasing the contacting area per stage (usually increases gas-side pressure drop),
(3) increasing scrubber liquor to gas ratio, and
(4) increasing lime or limestone utilization.

Some of these design measures not only affect SO_2 removal but also affect the scaling tendency of the system.

The most frequently used contactors for SO_2 removal are:

(1) venturi scrubbers,
(2) spray towers (horizontal and vertical),
(3) grid towers, and
(4) mobile bed absorbers (marble bed, TCA packed column).

The performance characteristics of each of these types of scrubbers are listed in Table 3-7.

Any of these scrubbers could be applied for both gas absorption and particulate removal, however packed columns show a much greater tendency for solids plugging. Resistance to plugging is an important parameter in scrubber selection. The open configuration of the spray tower gives it a lower gas-side pressure drop and makes it less susceptible to plugging when compared to the closed configuration of the marble bed and TCA scrubbers.

The volume of flue gas to be treated normally determines the physical size of the scrubbing device. The minimum and maximum velocities selected vary widely among the scrubber types but generally fall in a range of 1.5-7.6 m/sec (5-25 ft/sec). The highest gas velocities occur when using a venturi due to the small diameter of the venturi throat. These high velocities, however, must be decreased before the gas enters the process mist eliminators.

Absorber Operation

The amount of slurry circulated is critical. If the liquid to gas ratio is too low, the slurry will absorb too much SO_2 per volume and critical supersaturation will occur. Crystallization will then take place in the scrubber rather than in the reaction tank. The minimum L/G generally ranges between 4.7-14.7 liters/Nm^3 (35 and 110 gal/1000 scf), depending on inlet SO_2 concentration, type of scrubber, and lime or limestone reactivity. The lowest L/G's are used for venturi scrubbers and mobile bed absorbers such as marble beds. High L/G's are common in spray columns while TCA's generally use mid-range values. In general, L/G ratios are higher for limestone systems than they are for lime systems due to limestone's lower reactivity.

TABLE 3-7

COMPARISON OF SCRUBBER TYPES FOR A LIMESTONE WET SCRUBBING SYSTEM

Parameter	Scrubber Type				
	Marble Bed	TCA	Venturi	Grid Tower	Spray Tower
SO_2 Removal Efficiency	Good	Good	Fair	Good	Good
Particulate Removal Efficiency	Good	Good	Excellent	Good	Fair
Typical L/G (gal/1000 scf) for SO_2 Removal	40-70	50-80	20-50	50-100	70-110
Gas Side Pressure Drop (in H_2O) for L/G Above	8-12	6-12	8-20	1-7	1-3
Gas Velocity (ft/sec)	3-8	6-11	125-300	6-11	5-25
Dissolution of Solids	Good	Fair	Poor	Fair	Poor
Resistance to Solids Plugging	Fair	Good	Excellent	Fair	Excellent

Both calcium sulfite and calcium sulfate form scales. Calcium sulfate can form supersaturated solutions in the scrubber system. The rate of scaling is sensitive to the supersaturation of calcium sulfate. Test results from the TVA Shawnee test facility have shown that scrubber internals can be kept relatively free of scale if the sulfate (gypsum) saturation of the scrubber liquor is kept below about 135 percent (at 50°C). If supersaturation is unchecked, calcium sulfate dihydrate starts to crystallize on solid surfaces, forming a scale. The supersaturation can be controlled by seeding the liquid with calcium sulfate dihydrate crystals, which provide a large surface area on which the dissolved salts preferentially deposit.

Evidence has also been encountered that coprecipitation of calcium sulfite and sulfate may occur. This phenomenon may enable operation of the process in a mode where calcium sulfate concentration will not reach its normal saturation level and thus will not form sulfate scale. Operation in this mode seems to depend on the level of oxidation occurring in the system.

Changes in liquor pH can also cause scaling. The solubility of calcium sulfite decreases with increasing pH. If the pH is allowed to fall below 5, comparatively soluble calcium bisulfite is formed. With a subsequent increase in pH value, the bisulfite is converted to calcium sulfite which, being less soluble, crystallizes out and forms scale.

The pH of a freshly prepared limestone slurry is usually between eight and nine. On contact with the SO_2 in the flue gas the pH rapidly falls below seven, but below pH six the decrease in pH is slow until the slurry is exhausted. The efficiency of SO_2 removal is not appreciably affected until the pH drops below about 5.2. The effect appears to be independent of the type of limestone. Limestone scrubbers usually operate with an inlet and outlet pH range between 5.2-6.4. The scrubber pH can be changed by varying the limestone feed.

Mist Elimination

Mist eliminator operation has been a major trouble spot in lime/limestone scrubbing. All wet scrubbers introduce mist droplets into the gas, some more than others. The mist must be collected and separated to prevent corrosion and solid deposits on downstream equipment and to avoid high energy consumption in evaporating the mist in the gas reheater. Since the drops are relatively large, usually 40 microns and larger, they can be removed effectively by simple impingement devices, such as zig-zag baffles (chevrons) or cyclonic demisters. Practically all designers have used chevrons with the major exception being Detroit Edison, where a cyclonic vane-type eliminator was installed.

Chevrons have had trouble with both inefficient mist removal and with plugging by soft deposits and scale. Almost complete mist elimination by chevrons has been achieved by mounting them in a slanted or vertical position instead of the usual horizontal position so that the liquor can drain off. This prevents re-entrainment of the liquor in the gas. Plugging and scaling of mist eliminators can be prevented by washing with fresh water. Intermittent washing with a high pressure soot-blower type spray has been more successful than a lower pressure continuous wash. Wash trays and wet electrostatic precipitators have also been used as part of the mist elimination system. A wash tray is placed under a horizontal chevron to remove solids in the entrained mist and to collect wash liquor flowing off the chevron. Wet electrostatic precipitators remove both mist and residual dust in the flue gas.

Sludge Dewatering

The sludge dewatering step is used to concentrate the solids for ease of handling and disposal and to lower transportation costs. The clear liquor is usually recycled back to the process for reuse. Sludge dewatering methods consist of clarification, bed drying, centrifugation, vacuum filtration, and thermal drying. In addition to these methods, interim ponding is sometimes used as a dewatering procedure. The settling characteristics of the sludge determine the effectiveness of this technique.

Clarifiers are presently used in SO_2 removal systems as a primary dewatering device when the solids content of the sludge is low. Limestone scrubber sludges containing unreacted additive are reported to thicken well compared to lime sludges because of the coarse limestone present. Limestone processes sometimes produce a turbid supernatant liquor.

Because of the physical nature of sulfate crystals as opposed to sulfite, dewatering is improved by a higher sulfate/sulfite ratio. Therefore, good results (85-90 percent solids) have been reported for a sample obtained from the Chiyoda process, which produces in a sludge with an extremely high sulfate to sulfite ratio.

One engineering company currently markets an SO_2/ fly ash control process using a sludge dehydration operation after an alkali scrubbing system. Clarifier underflow at 30 percent solids concentration is raised to 90-95 percent solids by passing the slurry through a dehydrator co-current with the hot flue gases (300°F).

3.2.3 Typical Operations Relating to Sinter Plants

Lime/limestone wet scrubbing should require no new technical modifications for application to steel mill sinter plants. Sulfur oxide levels in the sinter plant effluent gas (25-1000 ppm) are in the same concentration range as effluent gas from oil and coal-fired boilers where the majority of the FGD systems have been installed. The oxygen content of the gas (12-16 mole percent) is about two to three times higher than that usually encountered from utility boilers. The higher oxygen concentration in the effluent gas should result in a high rate of calcium sulfite oxidation. This has proved to be the case from Japanese experience (see Appendix B) where 30-100 mole percent of the calcium sulfite was oxidized to calcium sulfate in the different FGD units.

The significant amounts of carbon monoxide (about 0.5-1.0 mole percent) should have no effect on the FGD system because of its relative stability. The water content of sinter plant gas generally ranges between 5-10 mole percent. The gas will evaporate large amounts of water from the initial liquid contacting device as is the case where FGD scrubbing is applied to utility boilers.

The particulate concentrations found in the gas are no higher than those commonly found on coal-fired boilers. High particulate removal efficiencies (over 95 percent) should be easily achieved by a particulate prescrubber such as a venturi.

The unburned hydrocarbons carried by the effluent gases are not normally encountered on oil or coal-fired utility boilers. A particulate scrubber will remove some of the hydrocarbons but the hydrocarbon removal efficiency by wet scrubbing has not been conclusively determined. Most of the hydrocarbons are aliphatic and will act as an inert in a wet scrubbing system. The addition of windbox gas recycle systems to sinter plants is expected to help control hydrocarbon emissions. It has been estimated that the recycle system at Wierton will reduce the concentration of unburned hydrocarbons in the effluent gases by 50 percent (CU-055).

3.3 Evaluation of USSR and Japanese Data

Data concerning FGD systems being used to treat waste gases from steel mill sinter plants in Russia and Japan were received and evaluated. The Soviet data for one limestone FGD system and the Japanese data for four FGD systems were summarized and examined for consistency with previous U.S. lime/limestone scrubbing experience. Technical notes describing these evaluations in detail, along with the actual Russian and Japanese data, are contained in Appendices A and B.

3.3.1 Summary of Russian Data

The Soviet data consisted of a description of the process and a discussion of operating parameters for the limestone scrubbing system applied to the Magnitogorsk sinter plant. The system removes 85 percent of the SO_2 and 50 weight percent of the particulates. A computer simulation was performed by using the given operating parameters and making several necessary assumptions to fill voids where information was lacking. Several results were found from the given data and the computer simulation.

(1) Oxidation was only 25 percent with a flue gas oxygen concentration of 15.9 percent.

(2) The SO_2 concentration in the flue gas was 1600 ppm. This is higher than the average concentration of 200 ppm believed to be found in U.S. sinter plant operations (ST-368, WO-092).

(3) The reported inlet and outlet particulate concentrations were 200 g/Nm^3 and 100 g/Nm^3, respectively. This is much higher than concentrations normally found on coal-fired boilers (4.6-16 g/m^3) or reported values for U.S. sinter plants (∿ 1 g/Nm^3). It was concluded that the particulate concentrations were probably incorrect due to misplaced decimal points. The actual values were probably 2 g/Nm^3 and 1 g/Nm^3 for the inlet and outlet flue gas particulate concentrations, respectively.

(4) The large particulate concentrations used in the process simulation model caused calcium sulfate to be subsaturated in the scrubbing system. The Soviets reported that 10 weight percent of the calcium value of the sludge was calcium sulfate.

(5) It was not considered worthwhile to make further calculations on the assumed error listed in (3) above.

3.3.2 Summary of Japanese Data

The Japanese data consist of process descriptions and discussions of operating variables for four FGD processes applied to treating waste gases from steel mill sinter plants in Japan. The processes are:

(1) Kawasaki Steel, Mitsubishi Heavy Industries (MHI) Process;

(2) Sumitomo Metal, Moretana Process;

(3) Kobe Steel Calcium Chloride (Cal) Process;

(4) Nippon Steel Slag (SSD) Process.

Two of the processes, the MHI and Moretana Processes, use a conventional lime or limestone absorbent. The Cal Process uses a lime absorbent in a 30 percent calcium chloride solution. The SSD Process uses a 40 weight percent CaO converter slag as an absorbent. A summary table showing all of the FGD processes that operate on Japanese steel mill sinter plants is included in the Japanese data found in Appendix B.

Several important findings were obtained from the Japanese data.

(1) The oxidation of calcium sulfite to sulfate in the scrubber was reported to be between 50-100 percent for the conventional lime/limestone absorbent processes.

(2) The SO_2 concentration in the flue gas fluctuated between 800-1200 ppm every 20 minutes. High and stable SO_2 removal was still obtained under these conditions.

(3) An inlet HCl concentration of 20-50 ppm in the flue gas was reported. High chlorine concentrations in wet scrubbing systems can cause corrosion.

(4) The oily matter in the incoming flue gas necessitated the use of oil resistant rubber linings to prevent swelling.

(5) SO_2 removal efficiencies of over 90 percent were reported for inlet SO_2 flue gas concentrations of 200-1200 ppm.

(6) Absorbent utilizations of 95 percent for lime and 80-85 percent for limestone were obtained.

Results from the evaluation of Russian and Japanese data were used to establish criteria for designing a limestone scrubbing system to treat waste gas from U.S. sinter plants. Section 4.0 describes, in detail, the approach used to design the limestone scrubbing system.

4.0 DESIGN APPROACH

Computer simulations were used to prepare conceptual process designs of limestone scrubbing systems for sinter plant applications. The Radian process model, a group of computer programs for simulating aqueous inorganic chemical processes, was used for these simulations. The process model performs calculations based on (1) chemical reaction rate and equilibrium calculations, and (2) process and equipment data which define the process flow scheme and characterize each of the individual process units. A thorough description of the Radian process model is provided in Appendix C.

Several information sources were utilized during this program to develop design data for input to the limestone process model. The open literature was screened to assemble available information. Mr. Richard Jablin was retained as a consultant and proved to be a valuable source of information. Data on a Russian sinter plant limestone process was obtained as a result of an EPA sponsored US/USSR technology interchange agreement. These data were evaluated to determine how best to incorporate the Soviet operating experience into a process design for U.S. sinter plants. A report describing Japanese technology for controlling sinter plant SO_2 emissions, prepared for Radian by Dr. Jumpei Ando, also provided valuable information.

Data from the above sources were evaluated to determine realistic design parameters for use as input to the computer model. A design basis for both a sinter plant and a limestone scrubbing system was determined. Section 4.1 describes in detail the design bases chosen for these systems.

Results of the conceptual designs were then used to size process equipment. An economic basis was selected and applied to the process designs to perform an evaluation of both capital investment and annual operating costs. A detailed description of the economic basis of this design is presented in Section 4.2.

4.1 Design Basis and Assumptions

A conceptual design basis for a limestone scrubbing system to remove SO_2 from steel mill sinter plant waste gases was developed from several sources. These included:

(1) an evaluation of U.S. steel mill sinter plant operations and typical emissions,

(2) a comprehensive technical data base developed by Radian in the area of lime/limestone wet scrubbing technology, and

(3) an evaluation of operating lime/limestone wet scrubbing systems in the USSR and Japan.

Data describing both sinter plant emissions and limestone scrubbing operations were evaluated to establish a realistic basis for the conceptual design. Sections 4.1.1 and 4.1.2 discuss the sinter plant and limestone scrubbing design parameters used as input to the computer model.

4.1.1 Sinter Plant Design Basis

The sinter plant design basis was developed from an EPA report describing the engineering and design of a sinter machine windbox gas recirculation system (PE-179). The No. 2 sinter machine of the Wierton Steel Division in Wierton, West Virginia, was the subject of that report. Two conceptual designs were developed from the Wierton report, one for standard sinter plant operation and one for a sinter plant with windbox gas recirculation. The recirculation of windbox effluent gas has the advantage of substantially reducing the volume of gas emitted from the sinter plant, thus reducing the amount of gas to be cleaned. In addition, recirculating a portion of the windbox effluent gas to the sinter strand allows unburned hydrocarbons in the gas to be combusted in the second pass through the sinter bed. The concentration of particulates in the effluent gas remains essentially unchanged but the amount of particulates leaving the system is reduced due to the reduced gas volume. These advantages make sinter machine windbox gas recirculation an attractive method for meeting future emission regulations. Therefore, a design of a scrubbing system for a plant with windbox gas recycle was performed in order to further identify the advantage of a gas recirculation system over a standard sinter plant operation.

The design basis chosen for the two plants is given in Table 4-1. Both sinter machines produce 6312 mtpd (6958 tpd) of sinter product. Sinter production in the U.S. ranges from 1350 to 9100 mtpd (1,500 to 10,000 tpd) for individual sinter strands (JA-136). The parameters for Case 1 (standard operation) are all taken from operating performance data for the Wierton No. 2 sinter machine except for the sulfur dioxide concentration in the gas. The actual SO_2 concentration in the gas from the sinter machine was less than 100 ppm. Since this concentration

TABLE 4-1
SINTER PLANT DESIGN BASIS*

Parameter		Case 1 (Standard Operation)	Case 2 (39% Windbox Gas Recycle)
Dry Gas		633,600 Nm^3/hr	386,640 Nm^3/hr
Total Gas		674,784 Nm^3/hr	422,640 Nm^3/hr
Gas Moisture Content		6.1 vol. %	8.5 vol. %
		Vol. %	Vol. %
Gas Composition (Dry)	N_2	77.1	73.3
	O_2	15.4	13.3
	CO	0.8	1.4
	CO_2	6.7	11.9
SO_2 Concentration (Dry)		750 ppm	1200 ppm
Particulates (Dry)		923 mg/Nm^3	923 mg/Nm^3
Condensible Hydrocarbons (Dry)		738 mg/Nm^3	369 mg/Nm^3
Gas Temperature		139°C (282°F)	219°C (426°F)

---

* Based on 6312 mtpd sinter product.

was already low, it was decided to assume an increased SO_2 concentration in the gas for design purposes.

Data from a U.S. sinter plant with an SO_2 concentration of approximately 900 ppm (YO-042) were checked to insure that the concentrations of other species in the gas such as N_2, O_2, and CO_2 would not be significantly changed by assuming a higher SO_2 concentration. No substantial change in concentration of the other species was found. Data from Soviet, Japanese, and other U.S. steel mill sinter plants corroborated this conclusion.

An SO_2 concentration of 750 ppm was chosen as the design basis for standard plant operations. Data from Soviet and Japanese sources indicated that an SO_2 concentration of 750 ppm represented an average value from processing high sulfur sinter mixes. Some sinter plants have concentrations that are at least that level (YO-042).

The design parameters for the second case (39 percent gas recycle on a dry basis) were taken from the design of a gas recirculation system for the same Wierton No. 2 sinter strand. The concentration of the unburned hydrocarbons was obtained from conversations with Wierton personnel (CU-055), and is a rough estimation.

The SO_2 concentration for the second case was calculated to be 1200 ppm (39 percent higher than for the standard sinter plant case). It was assumed that the amount of SO_2 evolved from the sinter strand would remain the same and that the SO_2 in the recycled gas would pass back through the sinter strand without further reaction. Therefore, the total amount of SO_2 evolved for both cases would be equal, although the SO_2 concentration of the gas evolved from the plant with windbox gas recirculation would be higher than for the standard operation case.

An average of the three particulate compositions presented in Table 3-3 was assumed for both cases. It was felt that an average composition would be representative of the wide variations in sinter plant particulate emissions. Emissions variations are caused by differences in sinter plant feed compositions. The particulate composition assumed for this study is given in Table 4-2.

The effect of varying particulate compositions on the operation and cost of a limestone scrubbing system should be insignificant. The overall system change resulting from the CaO and MgO content of the particulate material will be small since limestone is added to the pre-scrubber liquor loop to insure the desired SO_2 removal. The equipment in the pre-scrubbing section of the system is all plastic-lined to prevent corrosion. Therefore, differences in chloride content of the particulates will not affect the corrosion rate.

Oxidation of calcium sulfite to calcium sulfate is believed to be increased by certain catalysts in the particulates entering the system. Iron oxides are believed to be oxidation catalysts. The change of iron oxide content in the different particulate compositons could, therefore, change the amount of oxidation in the pre-scrubber and SO_2 absorber. A system designed for low oxidation could experience scaling problems if high oxidation was actually encountered. Also, limestone sludge with a high sulfate concentration dewaters more easily than a sludge with a low sulfate concentration. This would affect the size of the clarifer.

TABLE 4-2

COMPOSITION OF PARTICULATES IN SINTER PLANT FLUE GAS

Component	Weight Percent
Fe_2O_3	24.5
CaO	11.0
MgO	2.6
K_2O	4.6
SiO_2	3.9
Al_2O_3	3.1
Na_2O	0.6
ZnO	0.2
Chlorides	6.8
Sulfates	8.7
Hydrocarbons	14.8
Other	0.7
Loss on ignition	18.4
	100.00

The condensible hydrocarbon concentration used is in the normal range for steel mill sinter plants. The hydrocarbon concentration can vary to a large extent depending on the amount of hydrocarbons in the sinter feed. No reported data was found on the percentage of the hydrocarbons which are condensible versus non-condensible. The hydrocarbons in the sinter feed come primarily from oily turnings or from coke added as a fuel.

4.1.2 Limestone Scrubbing Design Basis

A limestone scrubbing system was selected instead of a lime scrubbing system for two reasons.

(1) Limestone systems are generally less expensive than lime systems, primarily because limestone is much cheaper than lime. Furthermore, lime prices are expected to escalate because of the cost of calcining limestone.

(2) Most steel plants typically use limestone in process operations and, as such, have a readily available supply.

It was necessary to select both limestone and make-up water compositions in order to simulate the limestone scrubbing system. The limestone composition presented in Table 4-3 was chosen based on previous pilot plant studies done by Radian. The make-up water composition listed in Table 4-4 was decided upon after conversations with Mr. R. Jablin (JA-137).

TABLE 4-3
LIMESTONE COMPOSITION

Component	Weight Percent
$CaCO_3$	97
$MgCO_3$	1
Inert	2

TABLE 4-4
MAKEUP WATER COMPOSITION

Parameter	mg/l
Total Solids	261
Total Dissolved Solids	238
Total Suspended Solids	23
Alkalinity as $CaCO_3$	76
$SO_4^=$	80
Cl^-	29
NO_3^-	2
HN_3	0.7
pH	7.2

Process Equipment

The following major equipment was included in the design to insure efficient FGD system operations. The criteria used for selecting each of these items is discussed below.

(1) Prescrubber

A venturi prescrubber was selected for sinter plant applications for three reasons.

(a) Flue Gas Cooling and Saturation - To avoid evaporation of water from the scrubber liquor and subsequent scale formation at the inlet to the SO_2 absorber, the flue gas should be presaturated.

(b) Chloride Removal - The high chloride concentrations in the flue gas would cause corrosion problems in the SO_2 absorption system if not removed.

(c) Particulate Removal - An additional benefit of using a venturi prescrubber is removal of most of the particulates entering the SO_2 scrubbing system.

(2) Forced Draft Fan

A fan is needed to overcome the pressure drop of the scrubbing system. The existing waste gas fan is designed to handle sinter plant flue gas at high temperatures (up to

175°C) and with a significant particulate loading (~ 1 g/Nm^3). No problems should be incurred in installing an additional forced draft fan to operate under these conditions. Induced draft fans, which are installed after the FGD system reheater have suffered from corrosion due to scrubber mist carry-over in the stack gas. Therefore, forced draft fans were selected for this design.

(3) Spray Tower SO_2 Absorber

A countercurrent vertical spray tower was chosen as the gas/liquid contacting device for SO_2 removal. This type of contactor was chosen over other types of contactors mentioned in Section 3.2. Spray towers have an open configuration which reduces the possibility of scaling and solids plugging in the scrubber.

(4) Ball Mill

A wet ball mill was placed in the FGD system to grind large calcium sulfite and sulfate crystals contained in a bleed stream of waste sludge from the clarifier bottoms. The ground crystals are recycled to the SO_2 absorber hold tank to insure that an adequate number of seed crystals are present to provide sites for calcium sulfite and sulfate precipitation. The formation of large sulfite and sulfate crystals in the system can cause a shortage of sites for precipitation. This can result in the relative saturation of calcium sulfite and sulfate in the liquor to rise above critical supersaturation values. The addition of seed crystal sites reduces the chances of scaling. Controlled seed crystal recirculation is a Radian proprietary concept.

(5) Demister

A vertical chevron demister located in a horizontal duct will be placed downstream of the SO_2 absorber. This type of demister has been shown to be effective in removing entrained mist without experiencing plugging problems. A high pressure mixture of fresh water and clarifier overflow will be used to intermittently wash the demister.

(6) Reheater

The reheater will be designed to heat the stack gas to 79.4°C (175°F) for stack gas buoyancy. A low pressure steam heat exchanger will be used to heat ambient air which will be blended with the stack gas stream to provide the necessary heat. This type of design will prevent fouling and corrosion problems experienced by reheaters placed in the stack gas duct. Low pressure steam needed for the reheater is available from steel mill operations.

Design Parameters

Design parameters used for the conceptual design of the limestone scrubbing system are given in Table 4-5. The limestone scrubbing basis is the same for both sinter plant applications to permit accurate comparisons.

Most sinter plants in the U.S. today use a high basicity sinter mix by adding limestone or dolomite to the sinter feed. The addition of the alkali causes some of the sulfur in the sinter bed to be fixed in the sinter as calcium sulfite and sulfate. It also produces a basic fly ash that contains 10-20 weight percent CaO plus MgO. These basic species will probably be soluble to some extent after removal from the gas stream in the prescrubber.

At one Soviet sinter plant facility, venturi scrubbers were used to capture fly ash which contained 10-13 percent CaO. They found that the CaO dissolved in the water to neutralize the acids present. The aqueous medium leaving the scrubbers was either weakly alkaline or neutral (SU-094). The venturi scrubbers removed up to 98.5 percent of the particulate and over 60 percent of the SO_2 (SU-093).

The venturi prescrubber system was designed to remove 30 percent of the SO_2 in the flue gas. This was done because the presence of basic fly ash components would already cause some SO_2 removal which would have to be considered in the overall system design. Also, removing part of the SO_2 in the venturi prescrubber would require less to be removed in the absorption section and would lower the absorption section equipment costs.

TABLE 4-5
LIMESTONE SCRUBBING DESIGN PARAMETERS

Design Area	Parameter	Design Specification (Weight Percent)
Venturi Pre-Scrubber	SO_2 Removal	30
	Particulate Removal	98
SO_2 Countercurrent Spray Scrubber	SO_2 Removal	90.4
	Particulate Removal	70
	Calcium Sulfite Oxidation	70*
Total System	Overall SO_2 Removal	93.3
	Overall Particulate Removal	99.4
	Overall Calcium Sulfite Oxidation	70*
	Solids in Scrubbing Slurry	12
	Solids in Clarifier Underflow	40
	Solids in Disposal Pond	60

*Mole Percent

Previous operating experience with venturi scrubbers, reported in the literature, indicated that a particulate removal efficiency of 98 percent can be easily obtained. Removal efficiencies of over 99 percent are not uncommon with venturi scrubbers. Therefore, the venturi prescrubber was designed to remove 98 percent of the incoming fly ash. Radian pilot plant experience indicated that an SO_2 spray scrubber can remove about 70 percent of the remaining particulates in the gas stream which had passed through an initial particulate collection device. The removal of 98 percent of the particulates in the prescrubber and 70 percent in the absorber results in an overall removal of 99.4 percent. The Radian conceptual designs have particulate concentations of 844-867 mg/Nm^3 at the inlet and 4.59-4.91 mg/Nm^3 at the outlet of the FGD System. This degree of particulate removal seems reasonable when compared to Japanese experience. The Moretana Process used at the Kashima Plant achieved greater than 90 percent removal with the use of two Moretana perforated plate scrubbers. The first scrubber is used to cool the gas and remove particulates while the second scrubber is used as an SO_2 absorber. The Radian design achieves a greater amount of particulate removal because the venturi prescrubber is capable of a much higher particulate removal efficiency than the perforated plate scrubber used at the Kashima Plant.

The conceptual design ot the two limestone scrubbing systems was based on removing an equivalent percentage of SO_2 from both systems while reducing the SO_2 level in the stack gases below 100 ppm. Equal SO_2 removal rates were used for both systems to achieve an equal basis for comparing the capital investment and operating costs. An overall SO_2 removal efficiency of 93.3 percent was chosen for both systems. This reduced the outlet SO_2 level to 44 ppm for the standard sinter plant case and 67 ppm for the recycle case. The emission rate for both particulates and SO_2 per unit of total strand feed is given in Table 4-6.

TABLE 4-6

MASS EMISSION RATES FROM THE RADIAN BASE CASE STEEL MILL SINTER PLANTS AFTER LIMESTONE SCRUBBING OF THE WINDBOX EXHAUST GAS

Basis: Mass emissions per unit of total strand feed including recycle fines and a hearth layer.

Strand feed rate = 336 mtph (370.4 tph)

Pollutant	Grams Pollutant / Kilogram Strand Feed	Pounds Pollutant / Ton Strand Feed
Particulates		
(a) Standard Case	0.011	0.021
(b) Recycle Case	0.006	0.013
SO_2		
(a) Standard Case	0.27	0.54
(b) Recycle Case	0.41	0.82

These SO_2 removal efficiencies are consistent with the SO_2 removal efficiencies experienced in Japan where greater than 90 percent SO_2 removal was achieved from scrubbing sinter plant gases with similar SO_2 concentrations. (See Japanese data in Appendix B)

Calcium sulfite and sulfate relative saturation levels in the scrubbing system are important parameters that affect the limestone system performance. It has been shown by testing at TVA's Shawnee pilot plant that sulfate relative saturation levels should be kept below a maximum of 1.35 to prevent scaling (EN-310). The maximum allowable relative saturation level for sulfite in lime/limestone systems is not as well defined. Relative saturation levels as high as six have been reported in systems which operated without scaling problems. It was decided to design the Radian scrubbing system to operate well below the maximum allowable saturation levels. Maximum sulfate and sulfite relative saturation levels of 1.16 and 2.25 in the scrubber bottoms were chosen to insure scale-free operation.

Data obtained from the open literature describing hydrocarbon removal from sinter plant operations is conflicting. Condensible hydrocarbon removal efficiencies are reported to vary from nil to almost 70 percent (ST-368, BA-444). The high removal efficiencies were obtained when using a high energy scrubber. From the available data, one cannot ascertain the degree of hydrocarbon removal to be expected from the venturi prescrubber included as part of the limestone system design. The hydrocarbons included in the particulate composition were treated as particulates and were the only hydrocarbons assumed to be removed by the scrubbing system. All hydrocarbons not collected in the scrubbers will exit the system with the stack gas from the absorber.

Limestone scrubbing systems normally operate using a limestone scrubbing slurry of 10-15 weight percent solids. A value of 12 weight percent solids was chosen for this design. The bleed stream from the SO_2 scrubbing system, which removes fly ash and calcium sulfite and sulfate, is normally thickened and sent to disposal. A clarifier was chosen as the thickening device for this system. The solids content of the clarifier underflow can range anywhere from 20 to 40 weight percent. An underflow concentration of 40 weight percent solids was chosen for this design because the solid waste is made up primarily of calcium sulfate dihydrate crystals which settle more easily than the calcium sulfite hemihydrate crystals. Most scrubber sludges contain more of the sulfite crystals than sulfate and do not settle as well.

The amount of oxidation of calcium sulfite to sulfate has been reported to be anywhere from 25-100 percent from Japanese and Soviet experience (see Appendix I and II). A value of 70 percent was chosen for the conceptual designs.

The clarifier underflow is transferred to a waste disposal site where the sludge settles to its final concentration. The high rate of sulfite oxidation in this system produces a sludge whose solids are composed predominantly of calcium sulfate crystals (about 65 percent of the solids). This relatively high concentration of sulfate crystals allows the sludge to compact to a final solids concentration of about 60 weight percent. Clear supernatant liquor from the disposal pond will be recirculated to the system.

4.2 Economic Basis

An economic evaluation of a limestone slurry flue gas desulfurization process requires that both total capital investment and annual operating costs be calculated. The basis for the economic calculations performed for this study was the January, 1975, report by McGlamery, et al. (MC-147) on detailed cost estimates for advanced effluent desulfurization processes. A direct comparison of the limestone scrubbing system design basis presented in Section 4.1 with the basis used by McGlamery allowed both equipment and operating cost estimates to be made. Cost estimates for equipment not included in McGlamery's report were based on work done by PEDCo (PE-146). The installed equipment cost was based on retrofitting the limestone slurry process equipment to an existing sinter plant.

4.2.1 Capital Investment Costs

From the design basis presented in Section 4.1, calculations were made to estimate the equipment sizes required to process flue gas from the two steel mill sinter plant operating cases considered. Once the equipment sizes for both cases were determined, size-cost scale factors presented by McGlamery, et al. (MC-147) were used to obtain estimates of the delivered equipment costs.

The criteria used to calculate the total capital investment required to install a limestone slurry flue gas desulfurization system on an existing steel mill sinter plant included the following.

(1) A cost index factor of 1.20375 (CH-278) was assumed in order to scale up the 1974 equipment costs presented by McGlamery (MC-147) to mid-1977 costs. This factor was calculated from the *Chemical Engineering* plant cost index based upon a September 1975 factor of 1.0514 (obtained by dividing the September 1975 index by the September 1974 index). An annual inflation rate of 7 percent was chosen for the years of 1976 and 1977.

(2) Installation costs for retrofitting the limestone slurry process equipment to existing sinter plants were based on cost estimate factors given by Wood (WO-078).

(3) The cost of spray tower SO_2 absorbers was based on data reported by the Western Precipitation Division of Joy Manufacturing Company (JO-194).

(4) The cost of the SO_2 absorber effluent clarifier was based on data presented by PEDCo (PE-146).

The items used in calculating the total capital investment for the limestone slurry process include both direct and indirect capital costs and are listed in Table 4-7.

TABLE 4-7

ITEMS USED TO ESTIMATE THE TOTAL CAPITAL INVESTMENT REQUIRED FOR A LIMESTONE SLURRY PROCESS

Direct Costs:

- Equipment (Purchased)
- Piping
- Structural Steel
- Concrete Foundations
- Insulation and Painting
- Electrical
- Instruments
- Buildings and Service
- Excavation, Site Preparation
- Auxiliaries
- Sludge Ponds (installed)

Fixed Costs (includes labor):

- Engineering Design and Supervision
- Construction Field Expense
- Contractor Fees
- Contingency

4.2.2 Annual Operating Costs

Table 4-8 lists both the direct and fixed operating costs which must be considered for an economic evaluation of a limestone slurry process. The estimates made for annual operating costs assumed a 1978 start-up of the limestone system. Operating cost data taken from several sources served as a basis for estimating 1978 costs (OT-043, MC-147, PE-146).

The estimates presented here for both operating and capital costs are much the same as those which might be calculated for comparable size power plant flue gas desulfurization units. No spare equipment for increased reliability has been included.

The results of the capital investment and annual operating cost estimates for both sinter plant operating cases are presented in Section 5.3.

TABLE 4-8

BREAKDOWN OF ANNUAL OPERATING COSTS FOR A LIMESTONE SLURRY PROCESS

Direct Costs:

Raw Materials
  Limestone

Conversion Costs
  Operating Labor and Supervision
  Utilities
  Maintenance
  Analyses

Fixed Costs:

Annual Capital Charges
(Includes depreciation, taxes, and insurance)

Overhead
  Plant
  Administrative

5.0 RESULTS

The results presented in this section reflect study efforts in two major areas:

(1) Preparation of conceptual process designs for the two sinter plant limestone scrubbing systems, and

(2) preparation of cost estimates, including both capital and operating costs, for the two limestone scrubbing system process designs.

Detailed discussions of the results of this study are reported in the following sections. Section 5.1 presents a description of the conceptual process designs for the two limestone scrubbing systems along with a detailed process flow diagram, material balances, and design specifications of the process components. Section 5.2 presents a plot plan of a potential arrangement for a limestone scrubbing system located in a sinter plant, and Section 5.3 contains the results of the economic evaluations of the two limestone scrubbing systems.

5.1 Process Designs

Design parameters presented in Section 4.1 were used as the basis for preparing conceptual process designs for limestone scrubbing systems to remove SO_2 from sinter sinter plant flue gas. Process designs were developed for both standard sinter plant and windbox gas recirculation operations.

A process flow diagram was prepared to depict the flow of process streams through the various equipment. Only one flow diagram was prepared since the limestone scrubbing system will employ the same types of equipment for both design cases. Figure 5-1 is the flow diagram illustrating the process stream flows in the limestone scrubbing system.

Two scrubber modules in parallel were used to treat the flue gas from the sinter plant while the limestone feed preparation, slurry processing, and calcium solids disposal sections of the process use only one set of equipment. However, for ease of depiction the process flow diagram shows only one set of equipment for the entire process.

Material balances were prepared for each design case to determine the process raw material requirements. In addition, material balance calculations were used to estimate the sizes of the various process equipment. Tables 5-1 and 5-2 present the results of the material balance calculations for the two design cases. The tables present the total process flow rates for both scrubbing modules.

The design specifications for the various process equipment were determined from the results of the material balance calculations. The function and operation of major equipment items are described, and their design specifications given in Table 5-3.

Potential problem areas in the FGD system were investigated. It was found that steel mill sinter plants normally operate 24 hours a day except for one eight-hour shift per week. The FGD system, during this period, would continue to circulate the scrubbing slurry to prevent settling of solids and plugging of the lines. Another potential problem area is the small amounts of contaminants contained in the flue gas. Species such as chlorides, fluorides,

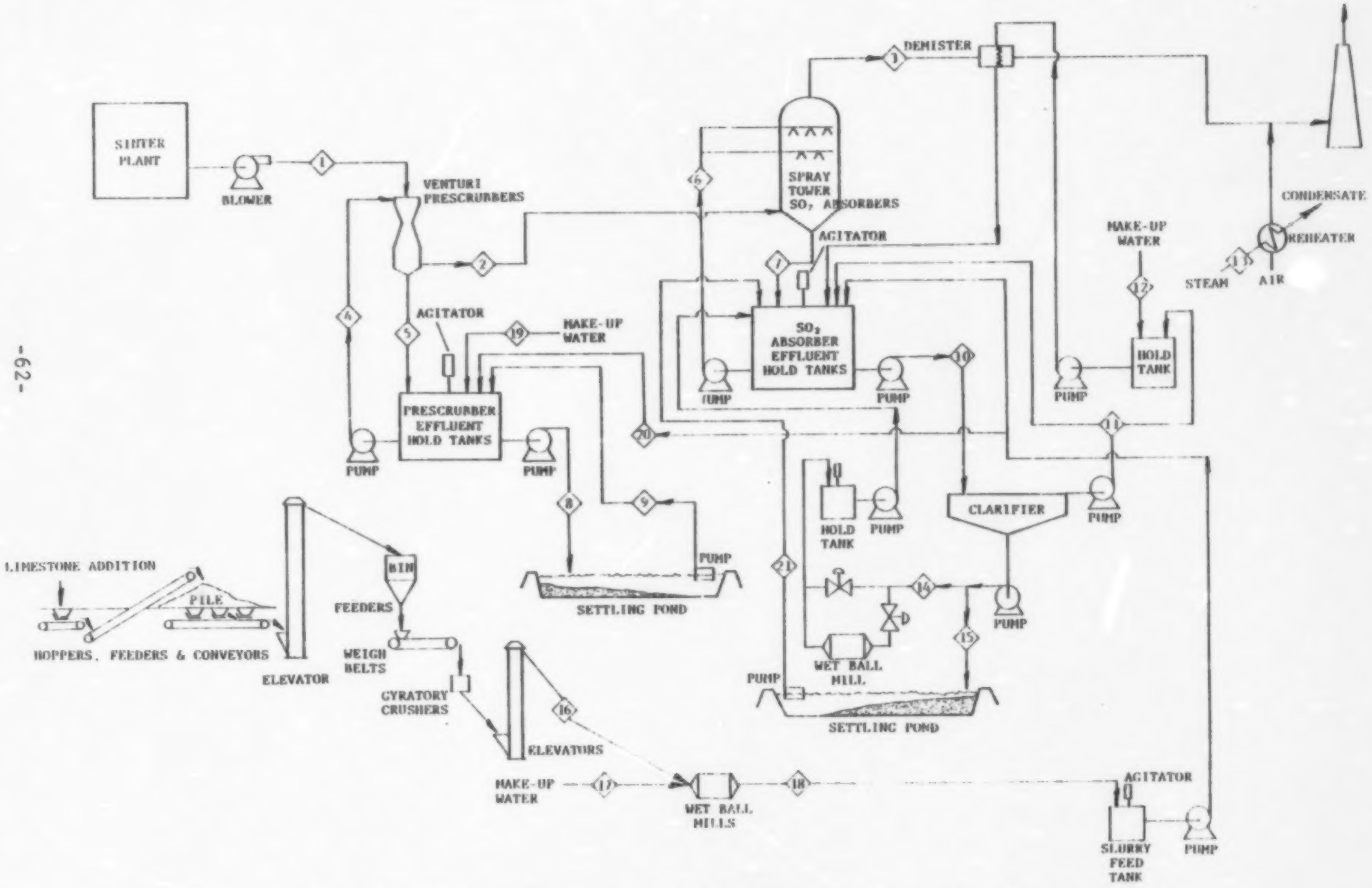

SINTER PLANT
BLOWER
1
VENTURI PRESCRUBBERS
2
4
5
AGITATOR
19
MAKE-UP WATER
PRESCRUBBER EFFLUENT HOLD TANKS
PUMP
PUMP
8
9
20
PUMP
SETTLING POND
3
DEMISTER
6
SPRAY TOWER SO₂ ABSORBERS
AGITATOR
7
SO₂ ABSORBER EFFLUENT HOLD TANKS
PUMP
PUMP
10
PUMP
HOLD TANK
MAKE-UP WATER
12
11
13
STEAM
AIR
REHEATER
CONDENSATE
HOLD TANK
PUMP
CLARIFIER
PUMP
21
14
PUMP
15
WET BALL MILL
PUMP
SETTLING POND
LIMESTONE ADDITION
PILE
HOPPERS, FEEDERS & CONVEYORS
ELEVATOR
BIN
FEEDERS
WEIGH BELTS
GYRATORY CRUSHERS
ELEVATORS
16
MAKE-UP WATER
17
WET BALL MILLS
18
AGITATOR
SLURRY FEED TANK
PUMP

FIGURE 5-1 PROCESS FLOW DIAGRAM - LIMESTONE SCRUBBING PROCESS FOR STEEL MILL SINTER PLANT APPLICATION

TABLE 5-1. MATERIAL BALANCE FOR A LIMESTONE SCRUBBING PROCESS ON A STANDARD SINTER PLANT

STREAM NUMBER →	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21
DESCRIPTION ↓	Gas to Prescrubber	Gas to SO_2 Absorber	Gas to Stack	Prescrubber Recycle	Prescrubber Bottoms	Absorber Recycle	Absorber Bottoms	Prescrubber Bleed to Pond	Prescrubber Pond Recycle	Absorber Bleed to Clarifier	Clarifier Overflow	Absorber Make Up Water	Steam Reheater	Clarifier Bottoms to Ball Mill	Clarifier Bottoms to Pond	Limestone Feed	Limestone Make Up Water	Limestone Slurry to Process	Prescrubber Make Up Water	Limestone to Prescrubber	Absorber Pond Recycle
Total Stream Flow Rate (g/sec)	243671	254053	253931	934789	924407	2617800	2618000	4075	3260	5836	4052	250.1	3646	17.4	1719	689.2	459.5	1148.7	10770	369.3	575.1
(g-mole/sec)	8370	8943	8941	46,333	45760	129780	129780	202	180.6	290	224	13.9	203	0.62	61.9	6.95	25.5	32.45	598	10.43	31.7
Temperature (°C)	139	50	50	50	50							21	126				21				
(°F)	282	122	122	122	122							70	259				70				
GAS PHASE																					
Gas Phase Flow Rate (g-mole/sec)	8363	8943	8941										202								
(Nm³/sec)	181.44	200.44	200.41																		
(g/sec)	245500	254050	253930										3645								
Composition (mole %)																					
N_2	72.4	67.7	67.7																		
O_2	14.5	13.5	13.5																		
CO_2	1.0	6.6	6.6																		
H_2O	6.1	12.2	12.2										100								
SO_2 (ppm)	104	461	44																		
LIQUID PHASE																					
Liquid Phase Flow Rate (g/sec)				822614	812073	2304200	2304400	3586	3260	5137	4052	250.1		10.4	1031.5		459.5	459.5	10770	147.7	575.1
(g-mole/sec)				45562	44987	127620	127620	198.6	180.6	285	224	13.9		0.58	57.1		25.5	25.5	598	8.20	31.7
Composition (molality $\times 10^{-4}$)																					
$SO_3^=$						0.435	0.666			0.435	.308	0		.308	.308		0		0		
HSO_3^-						1.81	4.21			1.81	1.20	0		1.20	1.20		0		0		
$SO_4^=$						116	120			116	109	7.11		109	109		7.11		7.11		
HCO_3^-						19.3	14.5			19.3	20.3	10.7		20.3	20.3		10.7		10.7		
H_2CO_3 (a)						10	11.5			10	9.83	1.76		9.83	9.83		1.76		1.76		
Ca^{++}						119	122			119	113	16.6		113	113		16.6		16.6		
$CaHCO_3^+$						2.21	1.69			2.21	2.22	0.229		2.22	2.22		0.229		0.229		
$CaSO_3$ (a)						3.27	5.06			3.27	2.25	0		2.25	2.25		0		0		
$CaSO_4$ (a)						61.4	10.2			61.4	62.0	1.21		62.0	62.0		1.21		1.21		
Mg^{++}						9.94	9.88			9.94	10.0	0		10.0	10.0		0		0		
$MgSO_4$ (a)						4.27	4.31			4.27	4.19	0		4.19	4.19		0		0		
Na^+						29.4	29.4			29.4	29.5	0		27.5	29.5		0		0		
Cl^-						34.9	34.9			34.9	34.8	14.08		34.8	34.8		8.08		8.08		
pH						6.48	6.29			6.48	6.51	7.12		6.51	6.51		7.12		7.12		
$CaSO_3$ Relative Saturation						1.453	2.250														
$CaSO_4$ Relative Saturation						1.089	1.134														
Dissolved Solids (wt ppm)						3044	3146			5044	2864	261		2864	2864		261		261		
Solids (wt %)				12	12.2	12	12	12	0	12				40	40			60		60	0
SOLID PHASE																					
Solid Phase Flow Rate (g/sec)	162.45	3.28	0.98	112175	112334	313620	313550	489		699				695	681.6	689.2		689.2		221.6	
(g-mole/sec)	2.250	0.0452	0.0136	771	773	2157	2155	3.36		4.81				0.05	4.75	6.95		6.95		2.23	
(mg/Nm³)	667	16.37	4.91																		
Composition (wt %)																					
CaO	11.0	11.0	11.0													0.97		0.97		0.97	
$CaCO_3$						11.85	11.74			11.85				11.81	11.81						
$CaSO_3 \cdot \frac{1}{2}H_2O$						21.01	21.02			21.01				20.99	20.99						
$CaSO_4 \cdot 2H_2O$						64.88	64.96			64.88				64.98	64.93						
$CaSO_4$	8.7	8.7	8.7																		
MgO	2.6	2.6	2.6																		
Na_2O	0.6	0.6	0.6																		
NaCl	6.8	6.8	6.8																		
Inerts	70.3	70.3	70.3			2.28	2.28			2.28				2.28	2.28	0.08		0.05		0.05	
Specific Gravity	3.15	3.15	3.15			2.42	2.42			2.42				2.42	2.42	2.73		1.65		1.65	

NOTE: All stream values are for the combined gas cleaning system.

TABLE 5-2. MATERIAL BALANCE FOR A LIMESTONE SCRUBBING PROCESS ON A SINTER PLANT WITH WINDBOX RECYCLE

STREAM NUMBER →	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21
DESCRIPTION ↓	Gas to Prescrubber	Gas to SO_2 Absorber	Gas to Stack	Prescrubber Recycle	Prescrubber Bottoms	Absorber Recycle	Absorber Bottoms	Prescrubber Bleed to Pond	Prescrubber Pond Recycle	Absorber Bleed to Clarifier	Clarifier Overflow	Absorber Make-Up Water	Steam Reheater	Clarifier Bottoms to Ball Mill	Clarifier Bottoms to Pond	Limestone Feed	Limestone Make-Up Water	Limestone Slurry to Process	Prescrubber Make-Up Water	Limestone to Prescrubber	Absorber Pond Recycle
Total Stream Flow Rate																					
(g/sec)	155233	165000	164890	583429	573662	1945800	1946000	3490.8	2792.6	5640	3948	243.6	1829.3	16.9	1675	672.4	448.3	1120.7	10049	360.5	558.3
(g-mole/sec)	5242	5780	5780	28919	28370	96469	96469	175.1	154.7	269.6	218.7	13.5	101.5	0.61	50.28	6.78	24.9	31.68	557.7	10.2	30.9
Temp (°C)	219	57	57										121								
(°F)	426	135	135										250								
GAS PHASE																					
Gas Phase Flow Rate																					
(g/sec)	155134	164990	164890										1829.3								
(g-mole/sec)	5241	5180	5780										101.5								
(Nm³/sec)	117.4	129.56	129.54																		
Composition (mole %)																					
N_2	67.1	60.7	60.7																		
O_2	12.2	11.0	11.0																		
CO_2	12.2	11.1	11.2																		
H_2O	8.5	17.1	17.1										100								
SO_2 (ppm)	1098	696	67																		
LIQUID PHASE																					
Liquid Phase Flow Rate																					
(g/sec)				513418	503554	1112700	1112900	3071.4	2792.6	4963	3948	243.6		10.1	1005		448.3	448.3	10049	144.2	558.3
(g-mole/sec)				28438	27889	94867	94869	172.2	154.7	264.9	218.7	13.5		0.56	55.68		24.9	24.9	557.7	8.0	30.9
Composition (molality × 10^3)																					
$SO_3^=$						0.390	0.426			0.390	0.294	0		0.294	0.294		0		0		
HSO_3^-						2.46	8.17			2.46	1.68	0		1.68	1.68		0		0		
$SO_4^=$						109.0	114.8			109.0	102.4	7.11		102.4	102.4		7.11		7.11		
HCO_3^-						18.49	7.53			18.49	19.98	10.7		19.98	19.98		10.7		10.7		
H_2CO_3(o)						13.38	16.46			13.38	13.0	1.76		13.0	13.0		1.76		1.76		
Ca^{++}						116.4	120.4			116.4	110.0	16.6		110.0	110.0		16.6		16.6		
$CaHCO_3^+$						2.18	0.909			2.18	2.26	0.229		2.26	2.26		0.229		0.229		
$CaSO_3$(o)						3.15	3.49			3.15	2.31	0		2.31	2.31		0		0		
$CaSO_4$(o)						70.55	75.21			70.55	64.90	1.21		64.90	64.90		1.21		1.21		
Mg^{++}						6.16	6.13			6.16	6.21	0		6.21	6.21		0		0		
$MgSO_4$(o)						2.64	2.71			2.64	2.59	0		2.59	2.59		0		0		
Na^+						18.25	18.26			18.25	18.26	0		18.26	18.26		0		0		
Cl^-						25.09	25.10			25.09	25.08	8.08		25.08	25.08		8.08		8.08		
pH						6.34	5.85			6.34	6.39	7.12		6.39	6.39		7.12		7.12		
$CaSO_3$ Relative Saturation						1.361	1.508														
$CaSO_4$ Relative Saturation						1.088	1.16														
Dissolved Solids (wt ppm)						2932	3099			2932	2757	261		2757	2757		261		261		
Solids (wt %)				12	12.2	12	12	12		12				40	40			60		60	
SOLID PHASE																					
Solid Phase Flow Rate																					
(g/sec)	90.13	1.98	0.594	70011	70108	833120	833050	418.9		676.8				6.8	670	672.4		672.4		216.3	
(g-mole/sec)	1.37	.0273	.00818	481	481	1602	1600	2.88		4.65				0.05	4.60	6.78		6.78		2.18	
(mg/Nm³)	844	15.28	4.59																		
Composition (wt %)																					
CaO	11.0	11.0	11.0																		
$CaCO_3$						11.8	11.7			11.8				11.8	11.8	0.97					
$CaSO_3 \cdot \frac{1}{2}H_2O$						21.0	21.0			21.0				21.0	21.0						
$CaSO_4 \cdot 2H_2O$						65.0	65.1			65.0				65.0	65.0						
$CaSO_4$	8.7	8.7	8.7																		
MgO	2.6	2.6	2.6																		
Na_2O	0.6	0.6	0.6																		
NaCl	6.8	6.8	6.8																		
Inerts	10.3	10.3	10.3			2.2	2.2			2.2				2.2	2.2	0.03					
Specific Gravity	3.15	3.15	3.15			2.42	2.42			2.42				2.42	2.42	2.15					

NOTE: All stream values are for the combined gas cleaning system.

TABLE 5.3

DESCRIPTION AND DESIGN SPECIFICATIONS FOR MAJOR PROCESS EQUIPMENT

Item	No.	Description	Equipment Design Parameters Parameter	Standard System	Recycle System
1. Blower	2	The blower is a forced draft fan that is installed prior to the FGD system. It is provided to overcome the FGD system pressure drop.	Power rating Gas flow rate Pressure drop	1530kw 141.4m^3/s 609mmH$_2$O	1145kw 105.8m^3/s 609mmH$_2$O
2. Prescrubber	2	The prescrubber is a venturi scrubber used to: 1. Cool and saturate the gas. 2. Remove particulates, hydrocarbons, chlorides, SO$_2$ and other undersirable emissions.	Gas flow rate Gas velocity	141.m^3/s 38.1m/sec.	105.8m^3/s 38.1m/sec.
3. Prescrubber Hold Tank	2	The prescrubber hold tank provides a location for limestone slurry addition to neutralize and react with absorbed compounds. Residence time is provided for calcium sulfite and sulfate precipitation.	Volume	174.1 m^3	199.4m^3
4. Absorber	2	The absorber is a countercurrent spray tower that is designed to remove a high percentage of the SO$_2$ and some of the remaining particulates.	Gas flow rate Gas velocity	118.6m^3/s 3.05m/sec.	78.4m^3/s 3.05m/sec.
5. Absorber Hold	2	The absorber hold tank is an agitated tank that provides a location for limestone addition to reach a constant slurry density. Residence time for sulfite and sulfate precipitation is provided.	Volume Liquid residence time	531m^3 7.2 minutes	708m^3 12.8 minutes
6. Demister	2	The demister is a vertical chevron demister installed in a horizontal duct after the spray tower. The demister removes entrained mist and solids in the stack gas.	Gas flow rate	118.5m^3/s	78.3m^3/s
7. Reheater	2	The reheater is a low pressure steam heat exchange that is placed outside of the stack duct. Air is forced through the exchange where it is heated and then mixed in with the stack gas. The stack gas is reheated to 79.4 for bouyancy	Power rating Steam requirement Surface area	4000 kw 1.83 kg/sec 146.3m^2	2000 kw 0.92 kg/sec 73.4 m
8. Clarifier	1	The clarifier is a solid-liquid separator that concentrates the solid waste to 40 weight percent solids. The clarifier overflow is low in suspended solids and is recycled back to the process.	Feed slurry flow rate	311 Liters/min.	303 Liters/min.

TABLE 5-3

DESCRIPTION AND DESIGN SPECIFICATIONS FOR MAJOR PROCESS EQUIPMENT

(Continued)

Item	No.	Description	Equipment Design Parameters Parameter	 Standard System	 Recycle System
9. Ball Mill	1	The ball mill is a continous wet ball mill that grinds a small slip-stream of the solid waste from the clarifier underflow. The outlet stream is sent back to the hold tank. The ball mill grinds the solid particles to provide additional surface area for calcium sulfite and sulfate precipitation in the hold tank. The ball mill will generally operate intermittently to control the quantity and size of sulfite and sulfate crystals.	Feed slurry flow rate	18.9 Liters/min.	18.9 Liters/min.
10. Prescrubber Settling Pond	1	The prescrubber settling pond is a disposal site for the particulates, calcium sulfite and sulfate sludge, and other component captured in the prescrubber. The waste sludge settles to a concentration of 60 weight percent solids. The pond is clay-lined and designed for a 40 foot depth.	Volume Solid waste feed rate Acres (@ 40 foot depth)	$403,050m^3$ 32 Liters/min. 8.2	$348,200m^3$ 27.6 Liters/min. 7.1
11. Absorber Settling Pond	1	The absorber settling pond is a disposal site for the particulates and calcium sulfite and sulfate sludge from the absorber. The waste sludge settles to a concentrated of 60 weight percent solids. The pond is clay-lined and designed for a 40 foot depth.	Volume Solid waste feed rate Acres (@ 40 foot depth)	$566,600m^3$ 45 Liters/min. 11.5	$554,700m^3$ 44 Liters/min. 11.2
12. Limestone Preparation System	1	The limestone preparation system crushes and grinds the limestone from 0 x 4 cm into 70 percent - 200 mesh. The limestone is slurried to a 60 weight percent solid slurry before being sent to the process.	Limestone feed rate	41.4kg/min.	40.3kg/min.

and arsenic have been reported to be present in sinter plant flue gas (VA-126, BA-449). These species can cause corrosion problems if allowed to build-up in the scrubbing system. These species will, however, be contained in the prescrubber loop where they will build up to some steady-state level. All corrosion problems can then be handled in the prescrubber loop and not affect absorber operations.

The FGD system designed for sinter plant applications includes an initial particulate collection system which is separate from the SO_2 absorption system. The systems are separated to prevent potential oxidation catalysts, chlorides, and other corrosion causing agents from entering the SO_2 absorption system.

Some important operating parameters for the two systems are given in Table 5-4. The major differences in the parameters for the standard and recycle cases are the liquid to gas ratio (L/G) and the liquid residence time in the hold tank. The recycle case has an L/G ratio that is 15.5 percent higher than for the standard case and the gas volume treated is about 35 percent less on a wet basis, thus the absorbed pumping requirements for the recycle case are about 25 percent less. Although the hold tank residence time is 78 percent higher for the recycle case, the actual required hold tank volume is only 33 percent larger because of the smaller liquid flow rate. Another difference in the two systems is an increased steam reheat requirement of 100 percent for the standard case due to both the increased volume of gas to be heated and the lower initial stack gas temperature.

The overall differences in stream flow rates and raw material requirements for the standard and recycle cases are small. This is due mainly to the fact that the same amount of SO_2 is being removed by both FGD systems. Major differences are evident, however, in the sizes of the gas handling equipment. These differences are reflected in the capital investment costs of the systems which are reported in Section 5.3.

TABLE 5-4

OPERATING PARAMETERS FOR PROCESS DESIGNS

Parameter	Standard Operation	39% Recycle
Design Gas Velocity (m/sec.)		
Venturi Prescrubber	38.1	38.1
Spray Tower SO_2 Absorber	3.05	3.05
L/G (Liters/Nm^3)		
Venturi Prescrubber	4.7	4.7
Spray Tower SO_2 Absorber	11.6(87.0)*	13.4(100.1)*
Design Pressure Drop (mmH_2O)		
Venturi Prescrubber	254	254
Spray Tower SO_2 Absorber	76	76
Ducting and Demisters	279	279
TOTAL	609	609
Liquid Residence Time (minutes)		
Prescrubber Effluent Hold Tank	7.6	12.0
SO_2 Scrubber Effluent Hold Tank	7.2	12.8
Solid Residence Time (days)		
Prescrubber Effluent Hold Tank	1.1	1.4
SO_2 Scrubber Effluent Hold Tank	9.3	12.6

*gallons/MSCF

5.2 Limestone Scrubbing System Layout

The conceptual process designs contained in Section 5.1 were used to prepare layouts of the limestone scrubbing systems. The layout, which is divided into two processing areas, was prepared to show the utilization of space for the process. The first area is the scrubbing section between the sinter plant and the stack which includes the scrubbers and hold tanks. This area is the most important in relation to space requirements in a retrofit situation because the area between the existing sinter plant and the stack is usually limited. Figures 5-2 and 5-3 show the layouts for the scrubbing sections of a limestone scrubbing system applied to standard and recycle steel mill sinter plant operations.

The second area is the limestone feed preparation and slurry processing section of the system. The layout for this section, shown in Figure 5-4, is the same for both sinter plant applications. This is because equal amounts of SO_2 are removed in both cases and, therefore, about the same amount of limestone is required and about the same amount of sludge is produced.

The equipment sizes were estimated from the process design data. A report by TVA was used as an aid in positioning some of the equipment on the layout.

The total space requirements for the two process designs are shown in Table 5-5. The recycle operation requires more land area for the scrubbing section because of the larger absorber hold tanks. The hold tanks are placed beneath the scrubbers in both designs. A length of 4.6m (15 feet) was chosen as a reasonable spacing between equipment in the scrubbing section.

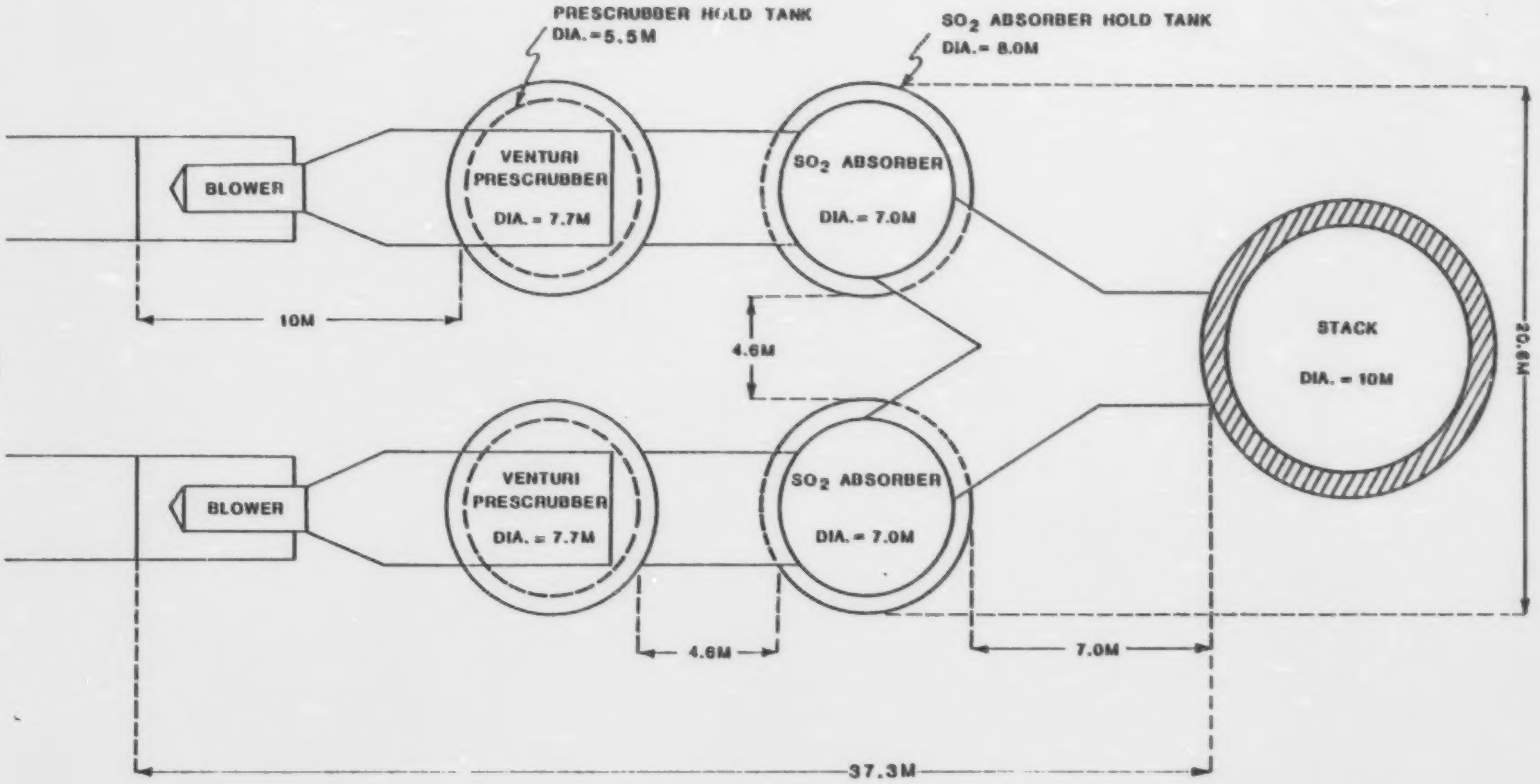

PRESCRUBBER HOLD TANK
DIA. = 5.5M
SO$_2$ ABSORBER HOLD TANK
DIA. = 8.0M
BLOWER
VENTURI PRESCRUBBER
DIA. = 7.7M
SO$_2$ ABSORBER
DIA. = 7.0M
STACK
DIA. = 10M
20.6M
10M
4.6M
BLOWER
VENTURI PRESCRUBBER
DIA. = 7.7M
SO$_2$ ABSORBER
DIA. = 7.0M
4.6M
7.0M
37.3M

FIGURE 5-2. LAYOUT OF SCRUBBING SECTION OF A LIMESTONE SCRUBBING PROCESS FOR A STANDARD STEEL MILL SINTER PLANT OPERATION

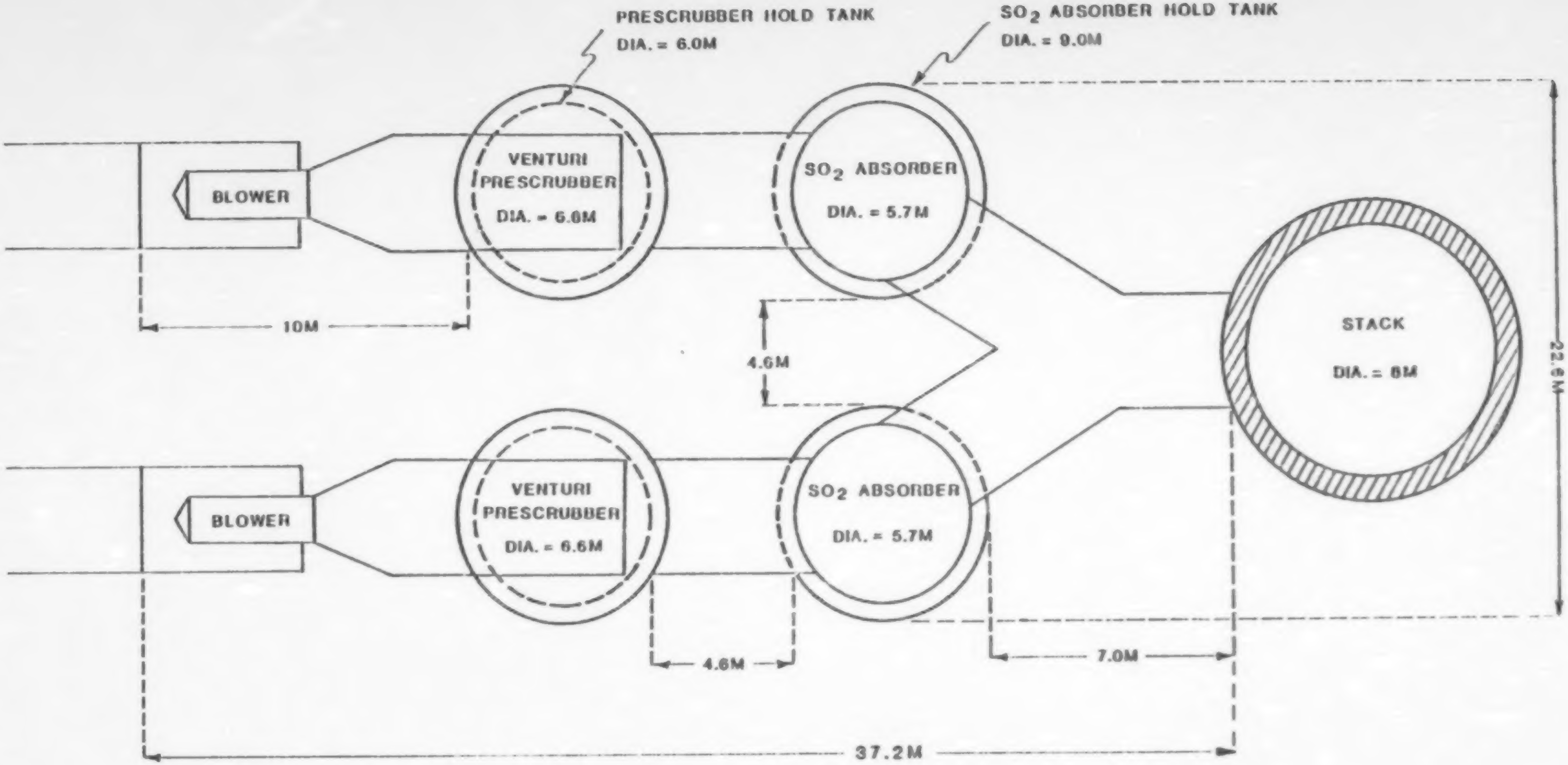
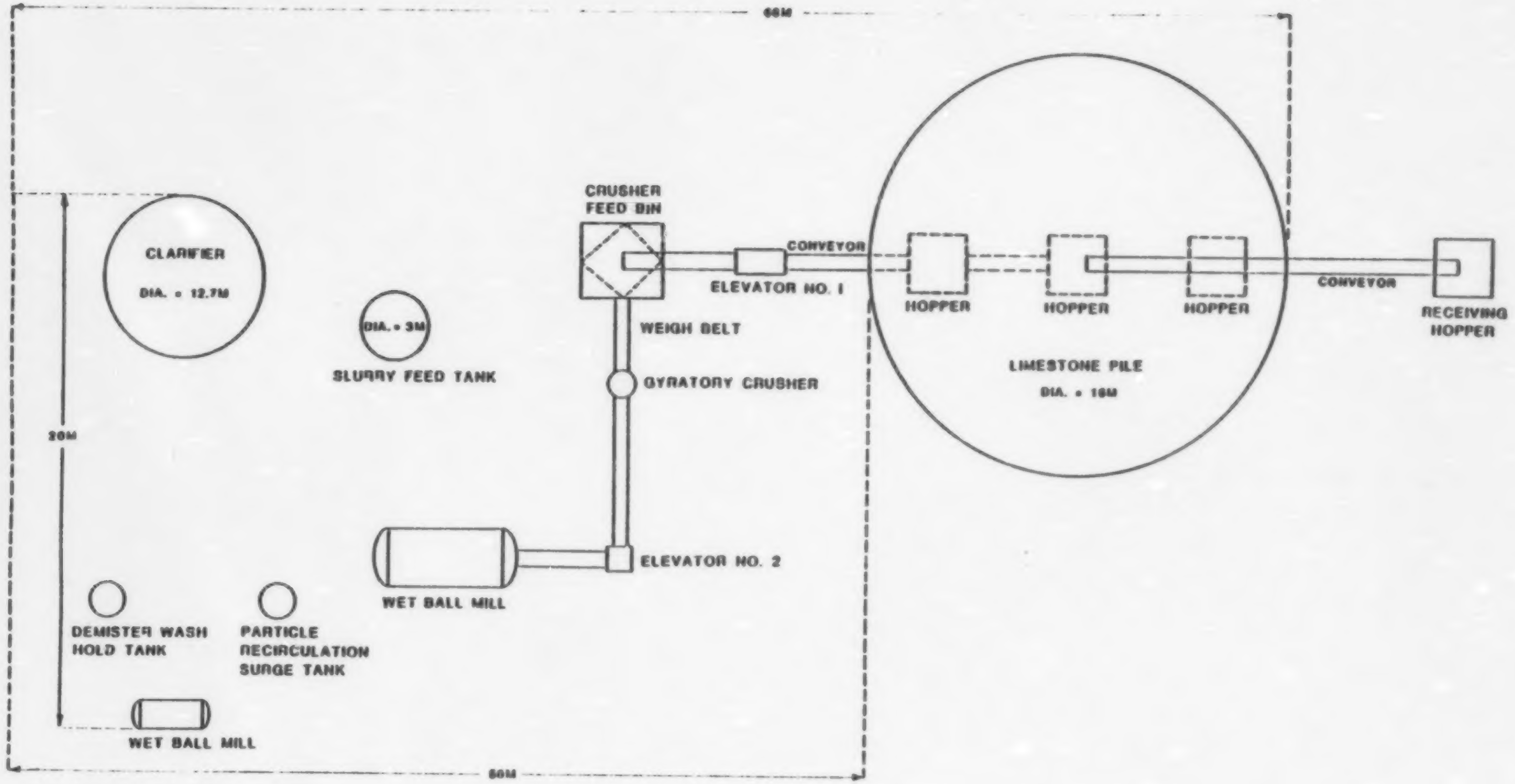

PRESCRUBBER HOLD TANK
DIA. = 6.0M
SO_2 ABSORBER HOLD TANK
DIA. = 9.0M
BLOWER
VENTURI PRESCRUBBER
DIA. = 6.6M
SO_2 ABSORBER
DIA. = 5.7M
STACK
DIA. = 6M
10M
4.6M
BLOWER
VENTURI PRESCRUBBER
DIA. = 6.6M
SO_2 ABSORBER
DIA. = 5.7M
4.6M
7.0M
37.2M
22.6M

FIGURE 5-3. LAYOUT OF SCRUBBING SECTION OF A LIMESTONE SCRUBBING PROCESS FOR A RECYCLE STEEL MILL SINTER PLANT OPERATION



66M
CRUSHER
FEED BIN
CONVEYOR
ELEVATOR NO. 1
HOPPER
HOPPER
HOPPER
CONVEYOR
RECEIVING
HOPPER
LIMESTONE PILE
DIA. = 18M
CLARIFIER
DIA. = 12.7M
DIA. = 3M
SLURRY FEED TANK
WEIGH BELT
GYRATORY CRUSHER
20M
ELEVATOR NO. 2
WET BALL MILL
DEMISTER WASH
HOLD TANK
PARTICLE
RECIRCULATION
SURGE TANK
WET BALL MILL
50M

FIGURE 5-4. LAYOUT OF FEED PREPARATION AND SLURRY PROCESSING SECTION OF A LIMESTONE SCRUBBING PROCESS FOR A STEEL MILL SINTER PLANT APPLICATION

TABLE 5-5

SPACE REQUIREMENTS FOR A LIMESTONE SCRUBBING SYSTEM ON STEEL MILL SINTER PLANT APPLICATIONS

Processing Area		Standard Operation	Recycle Operation
Scrubbing			
length (m)		37.3	37.2
width (m)		20.6	22.6
area (m^2)		768	841
Feed Preparation and Slurry Processing			
length (m)		66	66
width (m)		20	20
area (m^2)		1320	1320
Total Area (excluding waste disposal)	(m^2)	2088	2161
	(acres)	0.52	0.53
Prescrubber Sludge Settling Pond	(m^2)	33059	28560
	(acres)	8.2	7.1
Absorber Sludge Settling Pond	(m^2)	46473	45497
	(acres)	11.5	11.2
Total Area (including waste disposal)	(m^2)	81620	76218
	(acres)	20.2	18.8

For the prescrubbers, the hold tanks are smaller in diameter than the scrubbers so that the equipment spacing is measured from the outside diameter of the prescrubber. For the absorbers, the hold tanks are larger in diameter than the scrubbers so that the equipment spacing is measured from the outside diameter of the hold tanks. Since the absorber hold tanks for the recycle operation are larger than for the standard operation, the scrubbing area becomes larger.

As can be seen, most of the space required is for disposal of the fly ash and calcium solids. The FGD process itself requires only about 2270m^2 (24,400 ft^2). About 42 percent of this area or 950m^2 (10,200 ft^2) is needed for the scrubbing section between the sinter plant and the stack.

The most significant difference in land requirements between the standard and recycle sinter plant operations occurs in the waste disposal area. The recycle operation requires 5475 m^2 (59000 ft^2) less for particulate and calcium solids waste disposal. This difference is mainly due to the smaller prescrubber settling pond because of the lower particulate loading for the recycle case.

5.3 Economic Evaluation

The results of the economic evaluation performed for this study indicate that a reduction in both capital investment and annual operating costs of a limestone scrubbing system can be realized by employing windbox gas recirculation. For the cases considered here, a 21.3 percent and 23.2 percent reduction is indicated in capital investment and annual operating costs respectively. The cost of modifying the sinter plant for windbox recirculation was not considered in this evaluation.

The reduced gas volume and quantity of particulates to be removed were the major reasons for the lower required capital investment cost for the recirculation case. The 35 percent reduction in gas volume did not reduce the plant cost proportionately. This is typical for small-scale units where the initial fixed costs of the manufactured equipment represent a larger portion of the total investment than do the costs for larger equipment.

Annual operating costs for the gas recirculation case were lower because of both the reduced gas volume to be processed and the lower capital investment required. The reduced gas volume required less electricity for handling and less steam usage for reheating. The reduction in these two direct costs also resulted in a lower cost being assigned to plant overhead. The reduced capital investment resulted in both lower maintenance and annual capital charges.

The estimated total annual operating cost of the limestone scrubbing systems is \$2.07 per metric ton of product sinter for the standard operation case and \$1.59 per metric ton of product sinter for the windbox recirculation case. This results in a price increase of about 2 percent or \$5.00 per ton of product steel when 80 percent of the blast furnace charge is sinter (JA-145).

The detailed results of both the capital investment and annual operating cost estimates are presented in the following section.

5.3.1 Total Capital Investment

A comparison of the total capital investment costs required for limestone slurry scrubbing of the two sinter plant design cases considered is presented in Table 5-6. The major cost savings for Case B, the windbox recirculation system, are in the particulate scrubbing, SO_2 scrubbing, gas reheating, and gas handling areas. Capital cost reductions of 19.9, 34.5, 27.4, and 28.2 percent respectively occured in each of these areas because of both the reduced gas volume and quantity of particulates processed in the recirculation case. The lower values for installation and associated construction costs for the gas recirculation case reflect the lower cost for the process equipment in these four areas.

The reduction of the quantity of particulates to be scrubbed, due to gas recirculation, reduced the cost of particulate disposal by 13.3 percent. Both modes of operation require that the same quantity of SO_2 be removed from the gas stream. Thus, the cost of calcium solids removal is nearly identical in both cases.

As stated in Section 5.3, this evaluation indicates that the total capital investment required for flue gas desulfurization can be reduced by 21.3 percent if windbox gas recirculation is employed in the sinter plant. The capital costs associated with installing a windbox gas recirculation system were not evaluated for this study. However, these costs are discussed in the Wierton Report (PE-179). Additional work is being done under this program and definitive costs will be developed.

A detailed breakdown of the equipment costs for each of the limestone slurry scrubbing areas is given in Tables D-1 and D-2 of Appendix D.

TABLE 5-6

TOTAL CAPITAL INVESTMENT SUMMARY FOR STEEL MILL SINTER PLANT FLUE GAS DESULFURIZATION USING LIMESTONE SLURRY SCRUBBING

	Case A - Standard Operation		Case B - 39% Gas Recirculation	
Direct Costs:				
Process Equipment				
Materials Handling	$ 32,850		$ 36,210	
Feed Preparation	94,040		97,230	
Particulate Scrubbing	673,170		539,190	
SO_2 Scrubbing	1,026,800		672,600	
Gas Reheating	210,470		152,900	
Gas Handling	143,300		102,860	
Solids Disposal	178,200		191,800	
Services	128,570		128,570	
Particle Recirculation	31,950		31,950	
Subtotal		$2,519,350		$1,953,310
Equipment Installation		978,000		762,000
Piping		750,000		586,000
Structural Steel		125,000		100,000
Foundations		1,511,000		1,172,000
Insulation and painting		50,000		40,000
Electrical		176,000		137,000
Instruments		100,000		78,000
BL Building and Service*		126,000		98,000
Excavation and Fill Site Preparation		250,000		195,000
Auxiliaries		25,000		20,000
Sludge Ponds (installed)				
Particulate Disposal		120,000		104,000
Calcium Solids Disposal		170,000		166,000
Subtotal Direct Costs		6,900,350		$5,411,310
Indirect Costs:				
Engineering Design and Supervision		899,000		712,000
Construction Field Expense		955,000		756,000
Contractor Fees		506,000		400,000
Contingency		1,011,000		800,000
Subtotal Indirect Costs		3,371,000		$2,668,000
TOTAL CAPITAL INVESTMENT		$10,271,350		$8,079,310

*Battery Limit

5.3.2 Annual Operating Costs

The estimated annual operating costs for both design cases are summarized in Tables 5-7 and 5-8. Both the operating costs and capital charges are included in these tables. As can be seen from the values in these tables, the total operating cost for the limestone slurry scrubbing process can be expected to be reduced approximately 23.2 percent by windbox gas recirculation. This is due to the reduced utility requirements for gas handling and gas reheating and to the lower capital charges required when gas recirculation is used.

This cost reduction would be affected if capital charges and operating costs of the windbox gas recirculation system were included in the analysis. However, these costs were not available for inclusion in this study.

TABLE 5-7

LIMESTONE SLURRY PROCESS

TOTAL ANNUAL OPERATING COSTS

(Existing steel mill sinter plant - standard operation)

	Annual Quantity	Unit Cost, $	Total Annual Cost, $	Percent of Total Annual Operating Cost
Direct Costs				
Delivered raw material				
Limestone	17.4 M mtons	6.60/mton	114,600	3.01
Subtotal			114,600	3.01
Conversion costs				
Operating labor and supervision	18,000 man-hr	10.00/man-hr	180,000	4.73
Utilities				
Steam	89,500 M kg	3.31/M kg	296,000	7.78
Process water	290,000 M liters	0.029/M liter	8,400	.22
Electricity	23,270,000 kWh	0.028/kWh	651,600	17.12
Maintenance				
Labor and material, .09 x 6,900,350			621,000	16.31
Analyses			29,400	.77
Subtotal conversion costs			1,786,400	46.93
Subtotal direct costs			1,901,000	49.94
Indirect Costs				
Average capital charges at 14.9% of total capital investment			1,530,400	40.20
Overhead				
Plant, 20% of conversion costs			357,300	9.39
Administrative, 10% of operating labor			18,000	.47
Subtotal indirect costs			1,905,700	50.06
Total annual operating cost		2.07/mton product sinter	3,806,700	100.00

Basis:

Remaining life of sinter plant, 30 yr.

Stack gas reheat to 79.4°C.

Sinter unit on-stream time, 7,000 hr/yr.

Midwest plant location, 1978 operating costs.

Total capital investment, $10,312,550; subtotal direct investment, $6,898,550.

Sinter plant capacity of 6312 mtpd

TABLE 5-8
LIMESTONE SLURRY PROCESS
TOTAL ANNUAL OPERATING COSTS
(Existing steel mill sinter plant - 39% gas recycle)

	Annual Quantity	Unit Cost, $	Total Annual Cost, $	Percent of Total Annual Operating Cost
Direct Costs				
Delivered raw material				
Limestone	17.0 M mtons	6.60/mton	112,200	3.84
Subtotal			112,200	3.84
Conversion costs				
Operating labor and supervision	18,000 man-hr	10.00/man-hr	180,000	6.16
Utilities				
Steam	44,900 M kg	3.31/M kg	148,500	5.08
Process water	267,200 M liters	0.029 liters	7,800	.27
Electricity	16,800,000 kWh	0.028/kWh	470,400	16.10
Maintenance				
Labor and material, .09 x 5,411,310			487,000	16.67
Analyses			29,400	1.01
Subtotal conversion costs			1,323,100	45.28
Subtotal direct costs			1,435,300	49.12
Indirect Costs				
Average capital charges at 14.9% of total capital investment			1,204,000	41.21
Overhead				
Plant, 20% of conversion costs			264,600	9.06
Administrative, 10% of operating labor			18,000	.62
Subtotal indirect costs			1,486,600	50.88
Total annual operating cost		1.59/mton product sinter	2,921,900	100.00

Basis:

Remaining life of sinter plant, 30 yr.
Stack gas reheat to 79.4°C.
Sinter unit on-stream time, 7,000 hr/yr.
Midwest plant location, 1978 operating costs.
Total capital investment, $10,312,550; subtotal direct investment $6,898,550.
Sinter plant capacity of 6312 mtpd

6.0 CONCLUSIONS AND RECOMMENDATIONS

Conceptual process designs and cost estimates were performed for limestone scrubbing systems to control sinter plant emissions. The following conclusions and recommendations are presented to summarize the results of this effort.

6.1 Conclusions

1) Lime/limestone scrubbing technology has been demonstrated in Japan and the USSR to be an acceptable method for removing SO_2 from sinter plant effluent gases. Based upon foreign experience and the conceptual process design performed for this study, there is no apparent reason that lime/limestone scrubbing technology cannot be applied to controlling emissions from sinter plants in the United States. This study is, therefore, an initial assessment of the economic impact of lime/limestone scrubbing on domestic iron and steel plants.

2) The cost of sinter plant flue gas desulfurization was determined to be $2.07 per metric ton of product sinter for the standard operation case and $1.59 per metric ton of product sinter for the windbox gas recirculation case. This may increase the price of the steel product by about $5/ton. The effect of this price increase on the profitability of steel production needs to be evaluated for each specific installation.

3) Cost estimates indicate that windbox gas recirculation can reduce both capital investment and annual operating cost requirements for sinter plant limestone scrubbing units by approximately 20 percent. Estimates made during this study include only the costs associated with the limestone units and do not account for the additional cost of installing the windbox gas recirculation system.

4) Nearly all existing sinter plants have some type of particulate control following the cyclones which are the primary particulate collection device. The Radian conceptual design includes a venturi pre-scrubber to saturate and cool the hot sinter plant gas and to remove chlorides and particulates. This design would be applicable to an existing plant when the presently used particulate collection device following the cyclones was removed. To retrofit the SO_2 scrubbing system on a plant with an existing particulate scrubber, the pre-scrubber section would be excluded from the system design. It is estimated that excluding the pre-scrubber section of the design would decrease the overall cost by about 20 percent.

6.2 Recommendations

1) The results presented in Section 5.0 indicate that limestone slurry scrubbing is a technically

feasible method of flue gas desulfurization for steel mill sinter plants. The results also indicate that a complete limestone system for a sinter plant as described in Section 3.0 would cost about 8-10 million dollars. A recommendation of this study is that SO_2 removal in particulate scrubbers be investigated as an alternate SO_2 control technique for sinter plants. The purpose of such a demonstration would be to provide an acceptable technique for meeting SO_2 emission regulations while reducing the economic impact of flue gas desulfurization on sinter plant operation.

Two SO_2 scrubbing mediums that are recommended for consideration in this evaluation are: (1) limestone slurry and (2) a sinter fly ash slurry mixture.

Sinter fly ash usually contains between 10-15 percent CaO and MgO, two basic species capable of removing SO_2 in a wet scrubber. The basic species come from the limestone or dolomite that is mixed in with the sinter charge in order to produce a basic sinter. The largest consituents of the fly ash are usually iron oxides. Sinter fly ash entering the scrubber would be supplemented with fly ash removed in the initial particulate collection device as needed.

2) Theoretical, laboratory, and pilot plant investigations should be conducted to determine the

specific influences of sinter plant exit gas components on process operations, and to obtain data for process optimization. Of particular concern are the effects on sulfite oxidation and the potential corrosion effects in the prescrubber liquor loop.

7.0 REFERENCES

BA-444 Bayr, Robert B. and Richard J. Wachowiak, "Elimination of Hydrocarbon Emissions from the Sinter Plant", TMS-AIME Ironmaking Proc. 31, 55-58 (1972).

BA-449 Ball, D. F., A. F. Bradley, and A. Grieve, "Environmental Control in Iron Ore Sintering", Presented at the Minerals and the Environment Symposium, London, June 1974.

BA-450 Ban, Thomas Eugene, "Sintering Process", U.S. Patent 3,849, 115 (November 1974).

BA-451 Ban, Thomas E., "Process for Conditioning Sinter Draft for Electrostatic Precipitation of Particulate Material Therefrom", U.S. Patent 3,909, 189 (September 1975).

BA-452 Ban, Thomas E., "An Improved Sintering Process to Overcome Environmental Problems in the Sinter Plant", Preprint No. 75-B-6, Presented at the AIME Annual Meeting, New York, February 1975.

CH-278 "CE Plant Cost Index", Chem. Engr. 26 September 1975.

CU-055 Current, G. P., "Private Communication", National Steel Corporation, Weirton Steel Div., Weirton, West Virginia, April 1976.

EN-310 Environmental Protection Agency, Technology Transfer, Lime/limestone Wet-Scrubbing Test Results at the EPA Alkali Scrubbing Test Facility, 2nd Progress Report.

JA-136 Jablin, Richard, Private Communication, Richard Jablin and Associates, Blue Bell, Pa., 15 April 1976.

JA-137 Jablin, Richard, Private Communication, Richard Jablin and Associates, Blue Bell, Pa., 5 May 1976.

JA-145 Jablin, Richard, Private Communication, Richard Jablin and Associates Consulting Engineers, 25 August 1976.

JO-194 Joy Manufacturing Company, Western Precipitation Div., Basic Handbook of Air Pollution Control Equipment, Los Angeles, 1975.

MA-544 Manning, G. E. and F. E. Rower, "A Characterization of Air Pollutants from Sintering Plant Induced Draft Stacks", TMS-AIME Ironmaking Proc. 30, 452-460 (1971).

MC-147 McGlamery, G. G., et al., Detailed Cost Estimates for Advanced Effluent Desulfurization Processes, Final Report, EPA-600/2-75-006, Muscle Shoals, Alabama, TVA, January 1975.

MC-205 McGlamery, G. G., et al., "Flue Gas Desulfurization Economics", Presented at the 6th Flue Gas Desulfurization Symposium, New Orleans, La., March 1976.

OT-R-043 Ottmers, D. M., et al., Evaluation of Advanced Regenerable Flue Gas Desulfurization Processes, Draft Report, Radian Project No. 200-116, EPRI Contract No. RP 535-I, Austin, Tx., Radian Corporation, March 1975.

PE-146 PEDCo-Environmental Specialists, Inc., Flue Gas Desulfurization Process Cost Estimate, Preliminary Draft, Contract No. 68-01-3150, Task 2, Cincinnati, Ohio, May 1975.

PE-179 Pengidore, D. A., Sinter Plant Windbox Gas Recirculation System Demonstration, Phase I, Engineering and Design, EPA 600/2-75-014, Weirton, W. Va., Nat'l Steel Corp., August 1975.

PE-193 Petrov, Yu A., et al., "Cleanup of Flue Gases at a Sintering Plant", Metallurgist 15 (7), 420 (1970).

RO-219 Rowen, Harold E., "Protecting the Environment During Agglomeration", Presented at the Institute for Briquetting and Agglomeration, Hyannis, Massachusetts, August 1975.

RO-256 Rowen, Harold E., "Agglomeration - an Environmental Tool", in Proc. Inst. for Briquetting and Agglomeration, 13th Biennial Confernece, Colorado Springs, Colorado, August 1973, pp. 13 ff.

SC-328 Schmitt, R. J., "Corrosion Problems in a Sinter Plant Exhaust Gas System", Corrosion II, 425t - 29t (1961).

ST-368 Steiner, B. A., "Pilot Plant Testing of High-Energy Scrubbers for Sinter Plant Gas Cleaning", TMS-AIME Ironmaking Conf. Proc. 31, 59-69 (1972).

ST-398 Steiner, Bruce, Private Communication, Armco-Environmental, 27 April 1976.

SU-092 Suitlas, John R., "Emission Characteristics and Pilot Plant Studies on a Sintering Plant Windbox Discharge", TMS-AIME Ironmaking Proc. 30, 461-68 (1971).

SU-093 Suprunenko, R. S., et al., "Industrial Use of Venturi Pipes for Wet Cleaning of Sinter Gas", Metallurgist 12 (7), 358 (1967).

SU-094 Suprenko, R. S., et al., "Removal of Sulfur Dioxide from Sinter Gas", *Metallurgist 14* (9), 542 (1969).

SU-095 Suprenko, R. S., et al., "Wet Cleaning of Sintering Gas", *Metallurgist 9* (10), 539-41 (1964).

VA-003 Varga, John, Jr., et al., *A Systems Analysis Study of the Integrated Iron and Steel Industry*, PB 184 577, May 1969.

VA-126 Varga, John, Jr., *Control of Reclamation (Sinter) Plant Emissions Using Electrostatic Precipitators*, Final Report, EPA 600/2-76-002, Contract No. 68-02-1323, Task 32, Columbus, Ohio, Battelle-Columbus Labs., January 1976.

WO-078 Woods, Donald R., "Technique for the Estimation of Capital Costs for the Process Industry", Presented at the Symposium on Cost Estimation, Permian Basin Section of the AIchE, Odessa, Texas, April 1975.

WO-092 Woodward, Kenneth, Private Communication, EPA Emission Standards and Engineering Div., Durham, N.C., 26 May 1976.

YO-042 York Research Corporation, *Test Report of Sinter Plant Emissions at Bethlehem Steel Corp., Bethlehem Plant, Bethlehem, Pennsylvania*, Final Report, 3 volumes, Y-8479-18, Stamford, Connecticut, December 1975.

APPENDIX A

COMMENTS ON THE SOVIET DATA

1.0 INTRODUCTION

As a result of a technology interchange agreement between the U.S. and the USSR which encourages cooperation between the two countries in common areas of environmental concern, information describing the operation of a full scale Soviet limestone wet scrubbing facility was recently disclosed to the EPA. Since the scrubbing facility in question has been used to control SO_2 and particulate emissions from a USSR iron ore sintering unit, it was hoped that the Soviet data might provide a meaningful basis for the application of lime/limestone wet scrubbing technology to sintering operations in this country. Unfortunately, the Soviet data was not of sufficient quality to be of any real use in this area.

The comments which are presented in this report are based upon an engineering review of process data obtained from two sources (KH-027, LO-149). The specific objectives of the work which was performed as part of this assessment were:

1) to identify unique mechanical features of the Soviet sintering plant scrubbing process,

2) to determine normal operating ranges of important process variables, and

3) to identify potentially significant process problem areas.

The approach which was taken in order to accomplish these objectives involved both:

1) a thorough analysis of the available information to identify key process features and performance characteristics, and

2) a computer simulation of the system to check the reasonableness and consistency of the Russian data.

As a result of these activities, it was concluded that the Soviet data were lacking substantially in the areas of both completeness and accuracy. For this reason it does not appear that this information will lend any support to the limestone wet scrubbing process development efforts which are currently underway in this country. Justification for Radian's position in this regard is described in the next four sections of this report.

In Section 2, a general conceptual description of the Soviet scrubbing system and several important pieces of process performance data are presented. The data quality question is discussed in Section 3. Radian's approach to the development of a computer simulation of the Soviet system is described in Section 4. Simulation results are discussed in Section 5.

2.0 DESCRIPTION OF SCRUBBING FACILITY

The Magnitogorsk limestone scrubbing system consists of 21 scrubbers separated into 3 scrubber banks. Normally 16-17 scrubbers are operating with the remained down for cleaning. Figure 2-1 shows the process flow diagram for one scrubber system. Table 2-1 gives the operating parameters for one scrubber. Scrubbert inlet and outlet flue gas compositions are presented in Table 2-2.

The gas is first sent to a spray scrubber through a forced draft fan. Slurry recirculated from the hold tank scrubs the waste gases, removing 85 percent of the SO_2 and 50 wt percent of the particulates. The following equations give the overall SO_2 absorption reactions:

$$CaCO_3 + SO_2 \rightarrow CaSO_3 + CO_2 \tag{2-1}$$

$$CaSO_3 + \frac{1}{2}O_2 \rightarrow CaSO_4 \tag{2-2}$$

Spent scrubbing liquor is processed in a cyclone filter. The underflow from the filter is sent to a sludge tank prior to disposal. The overflow from the filter is sent to a hold tank where fresh limestone slurry (prepared in a limestone slurry tank) is added. The sludge tank and limestone slurry tank are common to 4-5 scrubbers.

A unique feature of this system is the location of the cyclone filter prior to the hold tank. Most lime/limestone scrubbing systems utilize a clarifier or filter after the hold tank to process a bleed stream from the recirculating slurry.

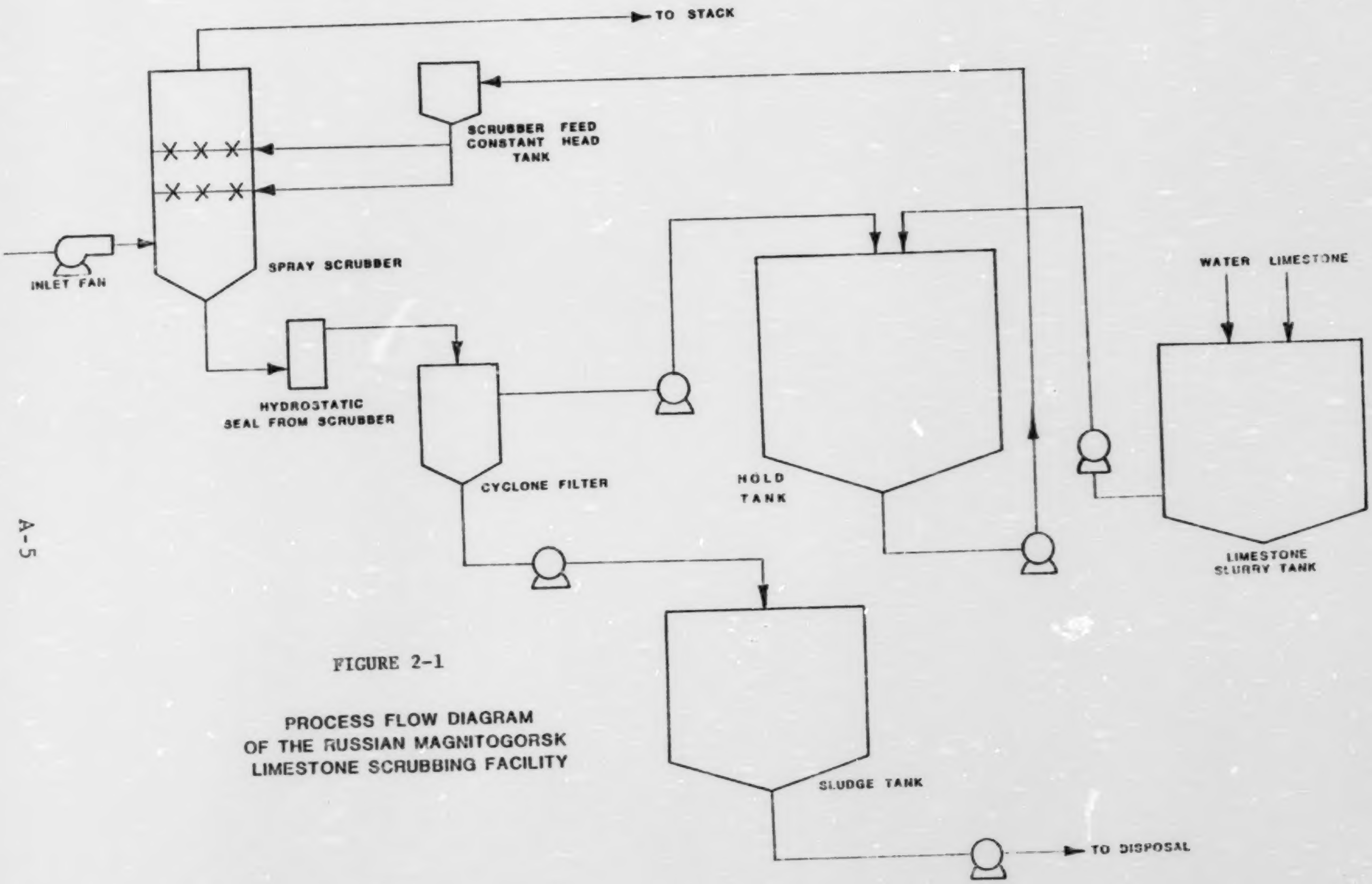

TO STACK
SCRUBBER FEED CONSTANT HEAD TANK
INLET FAN
SPRAY SCRUBBER
WATER
LIMESTONE
HYDROSTATIC SEAL FROM SCRUBBER
CYCLONE FILTER
HOLD TANK
LIMESTONE SLURRY TANK
SLUDGE TANK
TO DISPOSAL

FIGURE 2-1

PROCESS FLOW DIAGRAM
OF THE RUSSIAN MAGNITOGORSK
LIMESTONE SCRUBBING FACILITY

TABLE 2-1
OPERATING CONDITIONS FOR ONE SCRUBBER SYSTEM AT THE MAGNITOGORSK SCRUBBING INSTALLATION

Parameter	Value
Gas Flow Rate	200,000 Nm^3/hr
Liquid Flow	1,350,000 liters/hr
L/G	6.75 liters/Nm^3
Gas Velocity	2.5 m/sec
SO_2 Removal Efficiency	85%
Particulate Removal Efficiency	50%
Scrubber Inlet Slurry Density	5 wt %
Limestone Utilization	40%
Limestone Composition	87% $CaCO_3$ 2% $MgCO_3$ Balance - SiO_2
Waste Sludge Calcium Distribution	60% $CaCO_3$ 30% $CaSO_3$ 10% $CaSO_4$
Hold Tank Volume	180 m^3
Hold Tank Residence Time	8 minutes

TABLE 2-2

THE AVERAGE CHEMICAL COMPOSITION OF WASTE GAS BEFORE AND AFTER PURIFICATION

MAGNITOGORSK SCRUBBING FACILITY

System Location	Volume % (dry basis)*									
	CO_2	CO	O_2	N_2	CH_4	NO_x	SO_3	SO_2	Dust g/Nm^3	Average Gas Temperature (°C)
Scrubber Inlet	4.0	0.6	17.0	77.71	0.5	0.012	0.011	0.16	200	125
Scrubber Outlet	4.1	0.6	17.0	77.71	0.5	0.010	none	0.016	100	50

* Moisture content of gases (wet basis): inlet - 7.1 volume percent
outlet - 12.2 volume percent

3.0 DATA QUALITY

The data obtained for the Russian Magnitogorsk limestone scrubbing system would have to be characterized as incomplete. The primary areas where the data quality was questionable or non-existent were: 1) inlet and outlet particulate loadings, and 2) make-up water composition. A discussion of the problems associated with the inlet and outlet particulate loading data is included in Section 5.0 of this technical note. In summary, it appeared that the dust loading figures presented were two orders of magnitude too high.

The water make-up composition was not given in either of the sources which were reviewed. The composition of this stream is important because it determines to a great extent the concentration of the noncalcium dissolved salts in the system. Both the recirculating slurry and filter bottoms dissolved salt concentrations are affected by this parameter.

The other data that was given in the literature appeared to be reasonable for a limestone scrubbing system although somewhat incomplete. It would have been very helpful in verifying the results of the process simulation program to have had a total solids and liquid composition of one of the streams in the scrubbing system. The scrubber bottoms or hold tank slurry composition would have been preferred.

4.0 SYSTEM EVALUATION

The computer simulation of the Soviet scrubbing process was attempted in order to calculate important stream flow rates and compositions. The following assumptions for the system were made:

- The CO, N_2, CH_4, and NO_x species were assumed to be inert and to pass through the scrubber system.
- Vapor liquid equilibrium of CO_2 and H_2O occurs.
- The particulates collected in the scrubber were treated as insoluble inerts.
- The $MgCO_3$ and SiO_2 constituents of the limestone were treated as inerts.
- The make-up water dissolved salts composition was taken to be 1000 ppm NaCℓ.
- The cyclone filter underflow was taken to be 10 wt percent solids. The feed stream to the filter was 5 wt percent solids.
- The cyclone filter was assumed to operate at 75 wt percent overflow and 25 wt percent underflow for the total stream.

The computer model which was used to estimate steady-state operating conditions for the Soviet limestone scrubbing process was developed by Radian to simulate aqueous inorganic chemical processes. In addition to simulating the reactions in the system and the operation of each piece of equipment, a material balance around the entire system was generated by the computer. A discussion of additional assumptions that are incorporated in the computer model will be included in an appendix to the final report for this task. Since other subtasks will utilize the process simulation model, it was decided to present the discussion of the model in an appendix to avoid repetition in other sections of the final report.

5.0 DISCUSSION OF RESULTS

The pertinent results of the computer simulation are shown in Table 5-1. Two important results of the calculations can be seen. First, there is a high percentage of inerts (84-88 wt percent) in the solids in the system streams. Secondly, calcium sulfate was predicted to be at sub-saturated levels in all parts of the system.

The high percentage of inerts which were computed to be found in the process streams is explained by the high flue-gas particulate loading data (pickup = 100 g/Nm^3). This particulate loading is tremendously high as compared to other combustion operations such as coal-fired utility boilers which generate flue gases containing from 4.6-16 g/m^3 of particulate matter (prior to any particulate control device). Three possible reasons exist for the large number of particulates.

1) Cyclones or other particulate control devices were either not used prior to the SO_2 scrubber or the particulate control equipment was not adequately maintained.

2) The dust loading could include condensible and non-condensible hydrocarbons. These hydrocarbons are usually contained in the sinter mix feed due to borings, turnings, and other hydrocarbon laden materials from other parts of the steel mill.

TABLE 5-1

RESULTS OF THE SCRUBBER SYSTEM SIMULATION PROGRAM

	Solids Composition (wt %)			Relative Saturation		Partial Pressures (atm)		
System Location	$CaCO_3$	$CaSO_3 \cdot \frac{1}{2}H_2O$	Inerts	$CaSO_3$(s)	$CaSO_4$(s)	CO_2 ($x10^{-2}$)	SO_2 ($x10^{-6}$)	pH
Hold Tank	12.3	3.2	84.5	6.39	0.68	8.29	0.419	6.04
Scrubber Bottoms	8.9	3.3	87.8	0.98	0.81	7.97	10.57	4.91
Waste Sludge Stream	8.9	3.3	87.8	1.0	0.79	11.01	0.0141	6.36

3) The dust loading could be an incorrect figure. Experience in U.S. sinter plants using well maintained cyclones show particulate loadings in the waste gases on the order of 1-2 g/Nm^3. The loading reported by the Soviets would then be off by 2 orders of magnitude if their operation is consistent with U.S. experience. The fault in the data could then be attributed to misplaced decimal points.

It is concluded that this third possibility is the most reasonable explanation for the apparent inaccuracy of the Soviet data. Further evidence that helps to support this conclusion is the sub-saturation levels of the calcium sulfate in the computer simulation of the system. The sulfate saturation is low because of the high percentage of inerts in the system. The reported data shows that there is 10 wt percent $CaSO_4$ among the calcium species in the waste sludge. Calcium sulfate normally has to reach the saturation level to precipitate and be present in the waste sludge. The computer simulation for the process, therefore, did not agree with the reported behavior of calcium sulfate in the waste sludge. More information on the composition of the process streams is needed before the results of this process simulation case can be used to draw reliable conclusions.

If the inlet and outlet particulate loadings were reduced by two orders of magnitude to 2.0 g/Nm^3 and 1.0 g/Nm^3, respectively, then the calcium sulfate saturation level in the waste sludge would probably be reached and solid calcium sulfate would be precipitated. Further computer simulation using this assumption was not performed. Before further evaluation of the system is performed, it is recommended that more information be obtained from the USSR. The data already supplied should first be verified as to whether it is accurate. Secondly, more data on the operating parameters of the system should be obtained so that the process can be better characterized.

6.0 REFERENCES

KH-027 Kharkov, "Information on the development of a method of purifying agglomeration gases of sulfur dioxide". Point AM-1-2 of Appendix 3, First Working Meeting of the Soviet-American Branch Group for Ferrous Metallurgy on Methods of Preventing Atmospheric (Air) Pollution, pp. 1-5. Translated by E. Fitzback, SCI TRAN, Santa Barbara, Ca., 1975.

LO-149 Lowell, P.S., Trip Notes. Gaseous Emissions Abatement Project Visit to the USSR, 12-24, Oct., 1975. EPA Contract No. 68-02-1319, Task 38. Austin, Tx., Radian Corporation.

TR 76-261

INFORMATION ON THE DEVELOPMENT OF A METHOD OF PURIFYING AGGLOMERATION GASES OF SULFUR DIOXIDE

[Informatsiya o razrabotke sposoba ochistke aglomeratsionnydk gazov ot serniystogo angidrida]

Khar'kov

Point AM-1-2. of Appendix 3, First Working Meeting of the Soviet-American Branch Group for Ferrous Metallurgy on Methods of Preventing Atmospheric (Air) Pollution, pp. 1 - 5.

Translated for EPA by
SCITRAN (SCIENTIFIC TRANSLATION SERVICE)
SANTA BARBARA, CALIFORNIA

INFORMATION ON THE DEVELOPMENT OF A METHOD OF PURIFYING AGGLOMERATION GASES OF SULFUR DIOXIDE

Khar'kov

A Brief Description of the Technological Process for Trapping SO_2 from Agglomeration Gases.

In order to purify agglomeration gases of sulfur dioxide in the Soviet Union, a lime method has been widely used in fields of irrigated scrubbers, as the most effective and economical method. A line diagram of the device is shown in Figure 1. The flue gas extractor forces the agglomeration gases into the scrubber, where they are irrigated with a lime suspension by means of nozzles. The process of absorption and neutralization of SO_2 may be given in the form of a reaction:

$$CaCO_3 + SO_2 \rightarrow CaSO_3 + CO_2$$
$$2CaSO_3 + O_2 \rightarrow 2CaSO_4$$

The purified gas is discharged into the atmosphere through the smokestack. The constantly renewed lime suspension circulates in a closed system: part of the spent suspension is discharged into the slurry gutter and the required amount of fresh suspension enters the circulation collector.

Basic Technological Indices of the Industrial Device (for one system)

Amount of run-through gas	200,000 nm^3/hr
Flowrate of gas in scrubber	2.5 m/sec
Amount of irrigation in scrubber	1350 m^3/hr
Degree of purification of gas of SO_2	85%
Coefficient of efficiency	
Lime	45% - 50%

Mean intake concentration of SO_2 in the agglomeration gases	5 - 10 g/nm^3

Basic Equipment of Sulfur Purification Devices

The supply of agglomeration gas to the scrubber is carried out by the aid of a flue gas extractor with a productivity rating of 200,000 nm^3/hr,

head — 340 mm water column,
power of electric motor — 630 kw,
rpm — 735 rpm,

along a gas duct made of sheet steel. The gas duct is heat insulated with slag and coated with tin plate. The scrubber is hollow with a conical bottom and top; scrubber diameter - 6300 mm, scrubber height 24,000 mm. The scrubber walls are made of sheet steel and on the outside are heat insulated with slag and coated with sheet tin. They are faced on the inside with a diabase and acid resistant plate. The top of the scrubber is made of ordinary steel chemically protected with epoxide resin. Within the scrubber are mounted 3 round collectors with 2-inch involute nozzles. The collectors are made of stainless steel pipes 219 mm in diameter. The lime suspension is fed into the collector by a pump:

productivity — 1350 m^3/hr,
with a head of — 56 m water column,
power of electric motor — 500 kw, and
rpm — 735 rpm.

The impeller of the pump is made of a special alloy and the plate* is corundized. The pipes which supply the lime suspension are made of stainless steel. The operating lime suspension is fed to a cyclone filter from the scrubber via the hydraulic valve. The characteristics of the cyclone filter are: diameter - 2020 mm, height - 880 mm, width of the filter (*) aperture - 10 mm. (The remainder of this page is unreadable.)

---

* Translator's note: due to poor quality of foreign text, precise translation of this term is not possible.

The circulation collector is made of sheet steel and is faced on the inside with an acid resistant plate. From the circulating collector, the suspension proceeds through a pipe 500 mm in diameter to the intake pump. The spent suspension is drawn off through a rubberized slurry gutter into the slurry collector.

Diameter	--	4,250 mm
Height	--	3,670 mm
Volume	--	40 m^3

Transfer pumping of the spent suspension into the slurry removal system is carried out by means of pumps with a productivity rating of 600 m^3/hr,

a head of 36 mm water column,
electric motor power of 125 kw, and
rpm - 985 rpm.

Purified gas from the scrubber proceeds along a "pure" gas duct 4,000 mm in diameter faced on the inside with an acid resistant plate and proceeds to the smokestack, where it is discharged into the atmosphere. The smokestack 110 meters high is made of sheet steel covered on the inside with epoxide resin. Compensators in the "pure" gas ducts are made of rubberized fabric.

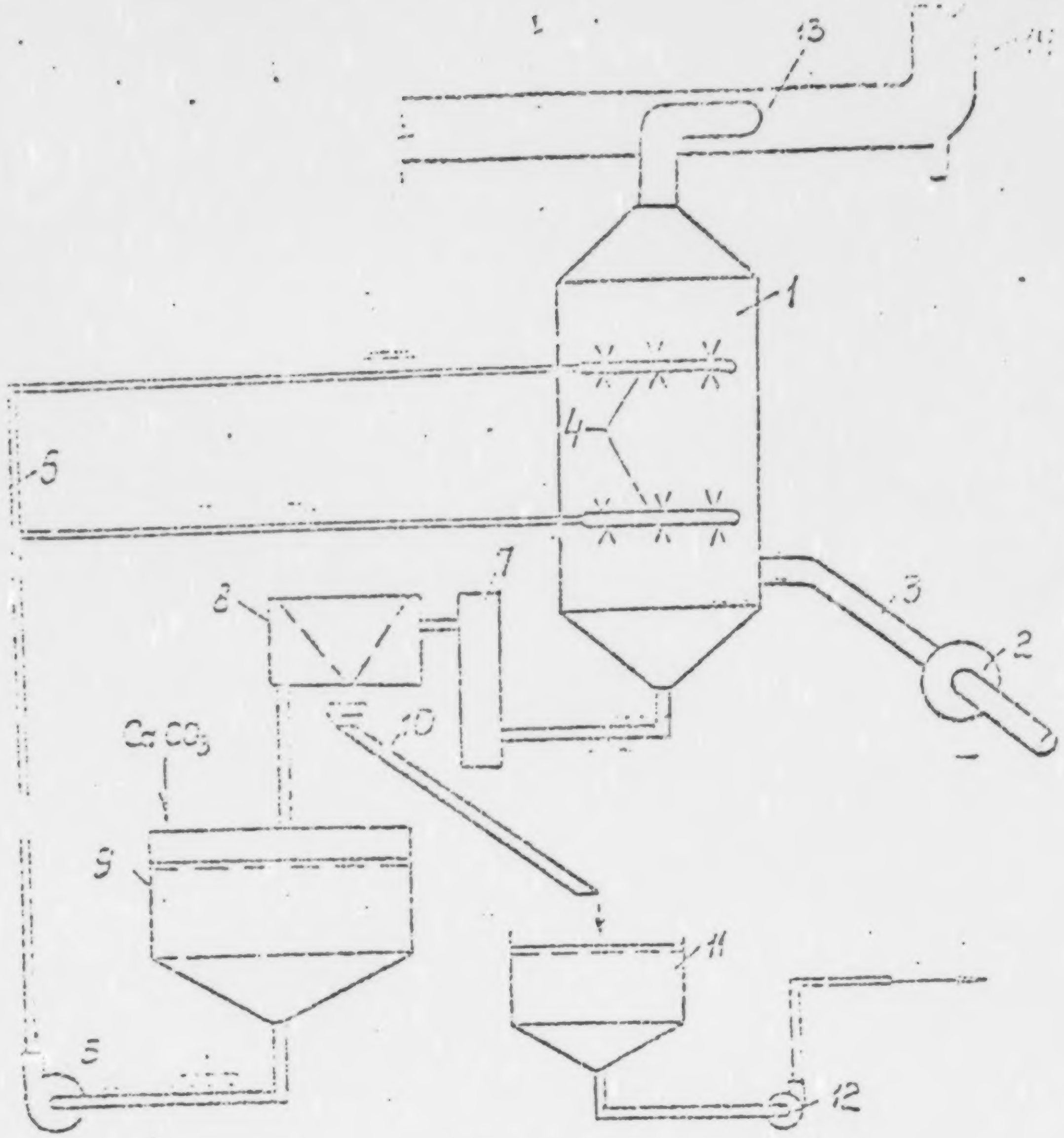
The Physico-Chemical Parameters of the Smoke-Gas Streams Before and After Purification.

The quantitative and qualitative composition of the agglomeration plant exhaust gases basically depends on the initial raw material and the conditions of agglomeration. Table 1 gives the average chemical composition of agglomeration gases before and after purification.

It is seen from the table that the agglomeration gases contain other components in addition to sulfur dioxide, for example, sulfur trioxide, carbon monoxide and nitrous oxide. Of the components of the agglomeration gases listed in the table, sulfur dioxide and sulfur trioxide as well as dust are primarily trapped. All of the other components are not practically absorbed by the lime suspension and are discharged into the atmosphere.

TABLE 1

Point of removing gas	CO_2 % vol.	CO % vol.	O_2 % vol.	N_2 % vol.	CH_4 % vol.	NO_x % vol.	SO_3 % vol.	SO_2 vol.	Dust g/nm^3	Average gas temperature °C
Before scrubber	4.0	0.6	17.0	77.71	0.5	0.012	0.011	0.15	200	125
After scrubber	4.1	0.6	17.0	77.71	0.5	0.010	none	0.016	100	50



13
14
1
4
5
3
2
8
7
10
CaCO3
9
11
5
12

Figure 1.

1- scrubber; 2- flue gas extractor; 3- dirty gas duct; 4- collectors with nozzles; 5- circulation pump; 6- supply pipe; 7- hydraulic valve; 8- filter; 9- circulation collector; 10- slurry gutter; 11- slurry collector; 12- slurry pump; 13- pure gas duct; 14- smokestack.

APPENDIX B

COMMENTS ON THE JAPANESE DATA

1.0 INTRODUCTION

The Environmental Protection Agency is interested in determining the technical and economic feasibility of applying lime/limestone scrubbing technology to control SO_2 emissions from steel mill sinter plants. Radian Corporation has been contracted by the EPA to investigate the feasibility of using lime/limestone scrubbing for this application. Since there has been no work in this area performed in the US, it was decided to examine the operating experiences of lime/limestone scrubbing units treating gases from steel mill sinter plants in other countries.

As part of a first step in the study, data from Dr. Jumpei Ando was obtained describing the operation of FGD scrubbing systems applied to Japanese steel mill sinter plants (AN-139). A copy of this information is included at the end of this Appendix. It was hoped that the Japanese data might provide meaningful information that would be helpful in the successful application of lime/limestone wet scrubbing systems to sinter operations in this country. Although some important process design information was lacking, the data provided information on the Japanese operating experience which should prove to be of considerable value.

The comments presented in this report are based upon a review of process data from Dr. Ando's report and four additional sources (AN-138, AN-141, PE-182, HI-149). The objectives of this assessment were:

(1) to identify unique mechanical features of Japanese sinter plant scrubbing processes.

(2) to summarize normal operating ranges of important process variables,

(3) to summarize Japanese operating experience, and

(4) to identify potentially significant process problem areas.

The approach taken to accomplish these objectives was to examine data for four wet scrubbing processes treating flue gas from Japanese steel mill sinter plants. The four processes investigated were:

(1) Mitsubishi Heavy Industries (MHI) Process installed at Kawasaki Steel plants (Lime is used as the absorbent);

(2) Moretana Process installed at Sumitomo Metal plants (Limestone is used as the absorbent);

(3) Kobe Steel Calcium Chloride Process (Cal Process) (A lime absorbent is used in a 30 percent calcium chloride solution);

(4) Nippon Steel Slag Process (SSD Process). (Converter slag containing 40 percent CaO is used as the absorbent).

In Section 2.0, a summary of the operating parameters and experience for each process is presented. A discussion of key process information is contained in Section 3.0.

2.0 PROCESS OPERATING EXPERIENCE

A description, including flow sheets, of the four processes investigated is contained in Dr. Ando's attached report. It was decided to summarize the operating parameters and experience in tabular form in order to be able to effectively determine important information. Table 2-1 contains a summary of the operating parameters for the MHI and Moretana process. The summary of operating parameters for the Kobe Steel calcium chloride process is given in Table 2-2. Insufficient data were presented to prepare a summary table for the SSD process.

It was hoped that enough information would be available for a computer simulation of the MHI and Moretana processes. Although a substantial amount of information was obtained from the different sources, key operating parameters for the processes which are necessary for a computer simulation were missing.

The Radian process simulation model requires the identification of process input streams and important operating variables. Information which was lacking in order to simulate the two Japanese processes were:

(1) limestone composition for both processes,

(2) water make-up composition for both processes,

(3) fly ash composition for both processes,

(4) the percent oxidation of calcium sulfite in the scrubber for the Mizushima plant (MHI process),

TABLE 2-1

SYSTEM DESCRIPTION OF THE MHI AND MORETANA PROCESSES

System Parameter	MHI Process General For All Sintering Plants	MHI Process Mizushima Plant (No. 4 Sinter Machine)	Moretana Process Kashima Plant
Absorbent	CaO	CaO	$CaCO_3$
Absorbent Characteristics		Lime has less than 1% MgO	Limestone is less than 1% MgO and is ground to pass 325 mesh
Absorbent Utilization	95 - 99	95	79 - 86
Gas Treated (1000 Nm^3/hr)	120 - 900	350 - 850 Designed for 750	350 - 880 Designed for 880
Type of Pre-Cooler	Spray Tower	Spray Tower	Moretana Scrubber
Pre-Cooler Dimensions (meters)		14 x 8, 26 high	6.5 Diameter, 24.5 high (2 coolers used)
Inlet Gas Temperature (°C)		57	
Outlet Gas Temperature From Cooler (°C)		150	150
Pre-Cooler Inlet Gas Composition (Vol. %)			
O_2	12 - 17	13.5	14 - 15
CO			1 - 1.5
CO_2	4 - 8		6 - 8
H_2O	6 - 12	5 - 13	4 - 10
Inlet SO_2 (ppm by Vol.)	600 - 1,200	400 - 1,100	200 - 450
Outlet SO_2 From Absorber (ppm by vol.)		20 - 50	3 - 15

TABLE 2-1 (Continued)

System Parameter	MHI Process General For All Sintering Plants	MHI Process Mizushima Plant (No. 4 Sinter Machine)	Moretana Process Kashima Plant
SO_2 Removal (%)	90 - 97	91 - 98	>95
Inlet Dust (g/Nm^3)	0.04 - 0.092 (ESP outlet)		0.15 - 0.23
Outlet Dust (g/Nm^3)			0.008 - 0.010
Dust Removal (%)			>90
Dust Components	Fe, Mn, Si, Pb, K, Na, Ca, Mg, Al, Zn, Cu, etc.		Ferric oxide
Inlet Cl (ppm as HCl)	20 -50		
Inlet Oily Matter (g/Nm^3)	0.057 g/Nm^3		
Type of SO_2 Absorber	Spray Tower	Spray Tower	Moretana Scrubber (perforated plate)
Absorber Dimensions (meters)		14 x 6.5, 30 high	6.5 Diameter, 20.5 high (2 absorbers used)
(1000 Nm^3/hr)			
Gas Velocity In Absorber (m/sec)		2.5	3 - 5
Circulating Liquor (m^3/hr)			2,500
L/G ($\frac{\text{Liters}}{\text{Nm}^3}$)		7	5
Slurry Conc. (wt %)	10 - 14		5 - 6
Inlet Liquor pH	6.5 - 7.5	6.4 - 7.5	6 - 6.5

TABLE 2-1 (Continued)

System Parameter	MHI Process General For All Sintering Plants	MHI Process Mizushima Plant (No. 4 Sinter Machine)	Moretana Process Kashima Plant
Oxidation In Scrubber (%)	70 - 100	Considerable Extent	50
Mist Eliminator Type			Vertical Chevrons In Horizontal Duct
Pressure Drop (mmH_2O)		Cooler, Absorber, and Mist Eliminator-120	Cooler - 220-250 Absorber - 120-140 Eliminator - 20-25 Total System - 700-800
Reheat Temperature (°C)		140	
References	HI-149	AN-139	AN-138, AN-139, AN-141

TABLE 2-2

SYSTEM DESCRIPTION OF THE KOBE STEEL CALCIUM CHLORIDE PROCESS

System Parameter	General For All Sinter Plants	Amagasaki Plant
Absorbent	$Ca(OH)_2$ in a 30% $CaCl_2$ solution	$Ca(OH)_2$ in a 30% CaCl2 solution
Absorbent Stoichiometry		1.05
Gas Treated (1000 Nm^3/hr)	50 - 375	350
Inlet Gas Temperature (°C)		120
Outlet Gas Temperature (°C)	70	70
Inlet O_2 Conc. (Vol. %)	Pilot Plant: 15 - 16	14 - 16
Inlet SO_2 Conc. (ppm)	Pilot Plant: 200 - 400	240 - 400
Outlet SO_2 Conc. From Absorber (ppm)		20
SO_2 Removal (%)	>90	91 - 94
Inlet Dust (g/Nm^3)		0.05 - 0.2
Dust Removal (%)	Most	50
Pre-Cooler L/G ($\frac{Liters}{Nm^3}$)		4
Slurry Conc. In Pre-Cooler (wt %)	5	
Type of Absorber	Spray Tower	Spray Tower

TABLE 2-2 (Continued)

System Parameter	General For All Sinter Plants	Amagasaki Plant
Absorber Capacity (1000 Nm^3/hr)		175 (2 absorbers used)
Absorber L/G ($\frac{Liters}{Nm^3}$)	2	3
Slurry Conc. In Absorber (wt %)	30	30
Absorber Inlet Liquor pH	7	6 - 8
Absorber Outlet Liquor pH	5.5	
Oxidation in Scrubber (%)	Pilot Plant - 30	750
Pressure Drop (mm H_2O)		Cooler, Absorber and Mist Eliminator - 190
References	AN-139, PE-182	AN-138, AN-139

(5) the scrubber effluent hold tank volume or residence time for both processes.

The last two pieces of information are most important, and are needed to adequately characterize any lime/limestone wet scrubbing system.

The operating experience of the four processes is given in Table 2-3. In general, the availability of the Mizushima plant (MHI process) and the two plants using the Moretana process was high.

TABLE 2-3

OPERATING EXPERIENCE OF LIME/LIMESTONE SCRUBBING

UNITS ON STEEL MILL SINTER PLANTS IN JAPAN

Parameter	MHI Process General For All Sintering Plants	 Mizushima Plant (Sinter Machine #4)	Moretana Process	Kobe Steel Cal Process	Nippon Steel SSD Process
FGD System	Absorbent-CaO Gas Treated-120,000-900,000 $\frac{NM^3}{hr}$	Absorbent-CaO Gas Treated-750,000 $\frac{NM^3}{hr}$	Absorbent-$CaCO_3$ Gas Treated- Kashima Plant-880,000 $\frac{NM^3}{hr}$ Wakayama Plant-370,000 $\frac{NM^3}{hr}$	Absorbent-$Ca(OH)_2$ in a 30 percent $CaCl_2$ solution Gas Treated- Pilot Plant-50,000 $\frac{NM^3}{hr}$ Amagasaki Plant-350,000 $\frac{NM^3}{hr}$	Absorbent-Converter Slag Gas Treated-Tobata Prototype Plant-200,000 $\frac{NM^3}{hr}$
Start-Up Date	First plant completed in 1973	November, 1974	Kashima Plant-September 1975 Wakayama Plant-May, 1975	Pilot Plant-June to December, 1974	1974
Monthly Availability (%)		90 - 95+	Kashima - 100 Wakayama - 98		
Operating Experience	1) The incoming gas laden with oily matter caused swelling of rubber linings of a pre-cooler on one installation. This problem was solved by using oil resistant rubber linings. 2) Using the pH of the recirculating slurry as a control factor, all of the scrubbers are running with excellent stability. The SO_2 concentration of flue gases from sintering plants changes	Several minor problems occurred in the first three months resulting in about 90% system availability: 1) Corrosion of the impeller of the cooler circulation pump. 2) Stop-up of the lime-slurry pump. 3) Breakage of a firebrick in the furnace(reheater). These problems were solved and nearly 100% availability	Wakayama Plant: The FGD system has been in smooth operation except for a defect in the plastic lining in a cooler which was found at the beginning of the operation and was repaired. Kashima Plant: Slight corrosion has been observed in the gas coolers of the FGD system which are constructed with 316 stainless steel and partially with a high Ni-Cr alloy.	Pilot Plant: 1) A soft deposit formed on the wall of the absorber when the L/G ratio was smaller than 1 but the deposit could be washed off by using an L/G larger than 2. 2) Corrosion has been the main problem. Stainless steel, plastic, and rubber linings are used for the material. Because	There has been some scaling problem to be solved to ensure a long-term continuous operation.

TABLE 2-3 (Continued)

Paramter	MHI Process		Moretana Process	Kobe Steel Cal Process	Nippon Steel SSD Process
	General For All Sintering Plants	Mizushima Plant (Sinter Machine #4)			
Operating Experience (Continued)	from 800 - 1,200 ppm every 20 minutes. Even at this kind of fluctuation, the FGD system operated with high and stable SO_2 removal. 3) The sulfite oxidation in the scrubber reaches as high as 100%. Scaling was prevented by: a) Addition of gypsum "seed" crystals to provide seed sites for calcium sulfate crystallization. b) High L/G was used. A scrubber with a simple structure was used to maintain uniform chemical conditions of the scrubbing slurry and to keep all the internals wet and clean. 4) Combustible flue gases from steel mills were burned to reheat the scrubber gas.	was obtained in the next three months.		the lower part of the cooler, where the hot gas is introduced was corroded severely, the part was replaced by titanium, which is durable. 3) Concentrations of magnesium and other impurities increased in the absorber liquor but caused no problem. 4) The plant operated without scaling and wastewater problems. Amagasaki Plant: The following problems were encountered during the two month test run: 1) Unusual vibration of a centrifuge. 2) Wearing of a control valve. 3) Scaling of a pH meter electrode. 4) Breakage of rubber lining in a reducer. These problems were solved and the plant went into commercial operation in April, 1976.	
References	HI-149	AN-139	AN-138, AN-139 AN-141	PE-182, AN-139	AN-139

3.0 DISCUSSION OF THE SYSTEM DATA

The operating parameters for the two lime/limestone processes (MHI and Moretana processes) are, of course, more important for this study since these systems are very similar to the type of FGD system considered in this project. But, the information obtained for the Cal and Slag processes should not be neglected since $CaCO_3$ and CaO are the absorbents in these systems respectively.

Several important pieces of information were obtained from the data on the MHI and Moretana processes.

(1) An inlet Cl concentration of 20-50 ppm in the flue gas was reported. High chlorine concentrations in wet scrubbing systems can cause corrosion.

(2) The oxidation in the scrubber was reported to be between 70-100 percent. Scaling was prevented by recycling gypsum as seed crystals, using a high liquid to gas ratio (L/G), and by using a scrubber with a simple structure.

(3) The oily matter in the incoming gas necessitated the use of oil resistant rubber linings to prevent swelling.

(4) The SO_2 concentration in the flue gas fluctuated between 800-1200 ppm every 20 minutes. High and stable SO_2 removal was still obtained under these conditions.

These four points need to be considered when designing a lime/limestone wet scrubbing system for a sinter plant facility.

It was reported for the Kobe Steel calcium chloride process that concentrations of magnesium and other impurities increased in the absorber liquor but caused no operating problems. The magnesium and other impurities mentioned did not cause a problem in this facility, but these species can affect process chemistry and should be considered in any FGD system design.

The Nippon Steel slag process experienced complete oxidation of the lime absorbent to calcium sulfate. It was reported that the prototype plant encountered some scaling problem that remains to be solved to insure a long-term continuous operation.

4.0 SUMMARY

Lime/limestone wet scrubbing has been successfully applied to steel mill sinter operations in Japan. About nine commercial lime/limestone scrubbing installations (excluding the Cal and SSD processes) will be operating on sinter plants by the end of 1976. Operating data from existing facilities have shown high system availabilities with minor operating problems. The SO_2 removal efficiency has been consistently over 90 percent with very high lime or limestone utilizations.

5.0 REFERENCES

AN-138 Ando, Jumpei, "Status of Flue Gas Desulfurization and Simultaneous Removal of SO_2 and NO_x in Japan", Presented at the Flue Gas Desulfurization Symposium, New Orleans, March 1976.

AN-139 Ando, Jumpei, "Desulfurization of Flue Gas from Iron-Ore Sintering Plants in Japan", Tokoyo, Chuo University, May 1976.

AN-141 Ando, Jumpei, Private communications, Chou University, 29 May 1976.

HI-149 Hirai, M., et al., "MHI Flue Gas Desulfurization Systems Applied to Several Emission Sources", Presented at the 6th Flue Gas Desulfurization Symposium, New Orleans, March 1976.

PE-182 Pedco Environmental, Inc., Untitled Preliminary Draft, regarding recent developments in desulfurization technology in Japan up to January 1975.

DESULFURIZATION OF FLUE GAS FROM
IRON-ORE SINTERING PLANTS
IN JAPAN

(May, 1976)

Jumpei Ando

Faculty of Science and Engineering

Chuo University

Tokyo

CONTENTS

1. INTRODUCTION
2. MIZUSHIMA PLANT, KAWASAKI STEEL (MHI PROCESS)
3. KASHIMA PLANT, SUMITOMO METAL (MORETANA PROCESS)
4. KOBE STEEL CALCIUM CHLORIDE PROCESS (CAL PROCESS)
5. NIPPON STEEL SLAG PROCESS (SSD PROCESS)
ADDRESSES OF STEEL PRODUCERS AND PROCESS DEVELOPERS

REMARKS

The metric system is used in this report. Some of the conversion figures between the metric and American systems are shown below:

1m (meter) = 3.3 feet
1m^3 (cubic meter) = 35.3 cubic feet
1t (metric ton) = 1.1 short tons
1kg (kilogram) = 2.2 pounds
1 liter = 0.26 gallon
1kl (kiloliter) = 6.29 barrels

The capacity of flue gas desulfurization plants is expressed in Nm^3/hr (normal cubic meters per hour),

1Nm^3/hr = 0.59 standard cubic foot per minute

L/G ratio (liquid/gas ratio) is expressed in liters/Nm^3.

1 liter/Nm^3 = 7.4 gallons/1,000 standard cubic feet

For monetary conversion, the exchange rate of 1 dollar = 300 yen is used.

1. INTRODUCTION

In 1974 the Japanese steel industry produced 114 million tons (metric tons) of crude steel and spent $563 million for pollution control, which was equivalent to 18.5% of the total investment made by the industry in that year. Many desulfurization units have been installed since 1971 to treat flue gas from iron-ore sintering plants, the major source of SO_2 emission from the steel industry (Table 1). As the absorbent, a lime slurry is used by Kawasaki Steel (MHI process), a limestone slurry by Sumitomo Metal (Sumitomo-Fujikasui Moretana process), a slurry of pulverized converter slag by Nippon Steel (SSD process) and a calcium chloride solution dissolving lime by Kobe Steel (Cal process). All of those plants by-produce gypsum. On the other hand, Nippon Kokan uses ammonia scrubbing to by-produce ammonium sulfate or gypsum by reacting lime with the sulfate.

By 1977, 22 FGD plants will be in operation with a total capacity of treating 13,800,000Nm^3/hr (8,120,000scfm) flue gas, which is about one-half the total gas from all sintering plants in Japan.

Flue gas from sintering plants is characterized by a high O_2 concentration (12-16%), relatively low SO_2 concentration (200-1,000ppm), and a dust content rich in ferric oxide. Oxidation of sulfite into sulfate occurs in the scrubbers much more readily than with flue gas from a boiler, because the oxidation is promoted by the high O_2/SO_2 ratio and also by the catalytic action of the ferric oxide.

The present report will describe mainly the lime and limestone processes and the dimensions and performance of the plants.

Table 1. SO_2 REMOVAL INSTALLATIONS FOR WASTE GAS FROM IRON-ORE SINTERING MACHINES

Steelmaker	Plant site	Gas treated (1,000Nm^3/hr)	Process	Absorbent	Year of completion	Gypsum (t/year)
Kawasaki Steel	Chiba	120	MHI	CaO	1973	3,600
"	"	320	"	"	1975	13,200
"	"	650	"	"	1976	26,500
"	Mizushima	750	"	"	1974	27,600
"	"	900	"	"	1975	32,400
"	"	750	"	"	1977	27,600
Sumitomo Metal	Kashima	880	Moretana	$CaCO_3$	1975	32,400
"	"	1,000	"	"	1976	40,500
"	"	1,000	"	"	1977	40,500
"	Wakayama	370	"	"	1975	14,400
"	Kokura	720	"	"	1976	26,500
Kobe Steel	Amagasaki	175 x 2	Cal	$Ca(OH)_2$	1976	12,600
"	Kobe	375	"	"	1976	12,600
"	Kakogawa	1,000 x 2	"	"	1977	72,000
Nakayama Steel	Osaka	375	"	"	1976	13,500
Nippon Steel	Tobata	200	SSD	Slag	1974	7,200
"	Wakamatsu	1,000	"	"	1976	32,400
Nippon Kokan	Keihin	150	NKK	NH_3, CaO	1971	7,200
"	Fukuyama	760	"	NH_3*	1976	12,000**
"	Ogishima	1,230	"	"	1977	20,000**

* Ammonia in coke oven gas ** Ammonium sulfate

2. MIZUSHIMA PLANT, KAWASAKI STEEL (MHI PROCESS)

Layout of Sintering Machines and FGD Plants

Kawasaki Steel installed FGD plants first at its Chiba works using the lime-gypsum process developed by Mitsubishi Heavy Industries. Satisfied with their operation, Kawasaki Steel introduced larger FGD plants at its Mizushima Works, where four iron-ore sintering machines with a unit capacity of 8,000-15,000 tons/day had been installed. The FGD plants were erected in an iron ore storing yard adjacent to the sintering plants (Figure 1). FGD plants for No. 3 and No. 4 sintering machines are in operation and a plant for No. 1 and No. 2 machines is under construction.

No. 4 Sintering Machine

Specifications of the No. 4 machine are shown in Table 2. The machine has 22 wind boxes. The amount, temperature and composition of gas from the wind boxes are shown in Figure 2. The total amount of the gas reaches 1,100,000Nm^3/hr. The SO_2-rich gas, about a half of the total, is selected by means of dampers as is shown in Figure 3 and sent to the FGD plant. The SO_2 concentration in the selected gas ranges from 400 to 1,100ppm while that in the rest of the gas ranges from 40 to 90ppm.

Table 2. SPECIFICATIONS OF No. 4 SINTERING MACHINE

Equipment	Type	Specifications
Sintering machine	Dwight-Lloyd	Capacity 15,000t/day Fire grate 410m^2
Main blower	Induced fan	Capacity 21,000m^3/min x 2 Motor 7,800kW x 2
Dust collector	Electrostatic precipitator	Capacity 21,000m^3/min x 2 Dust Inlet 1.0g/Nm^3 Dust Outlet 0.06g/Nm^3
Cooler	Lurgi	Cooling area 550m^2 Cooling fan 15,000m^3/min x 2

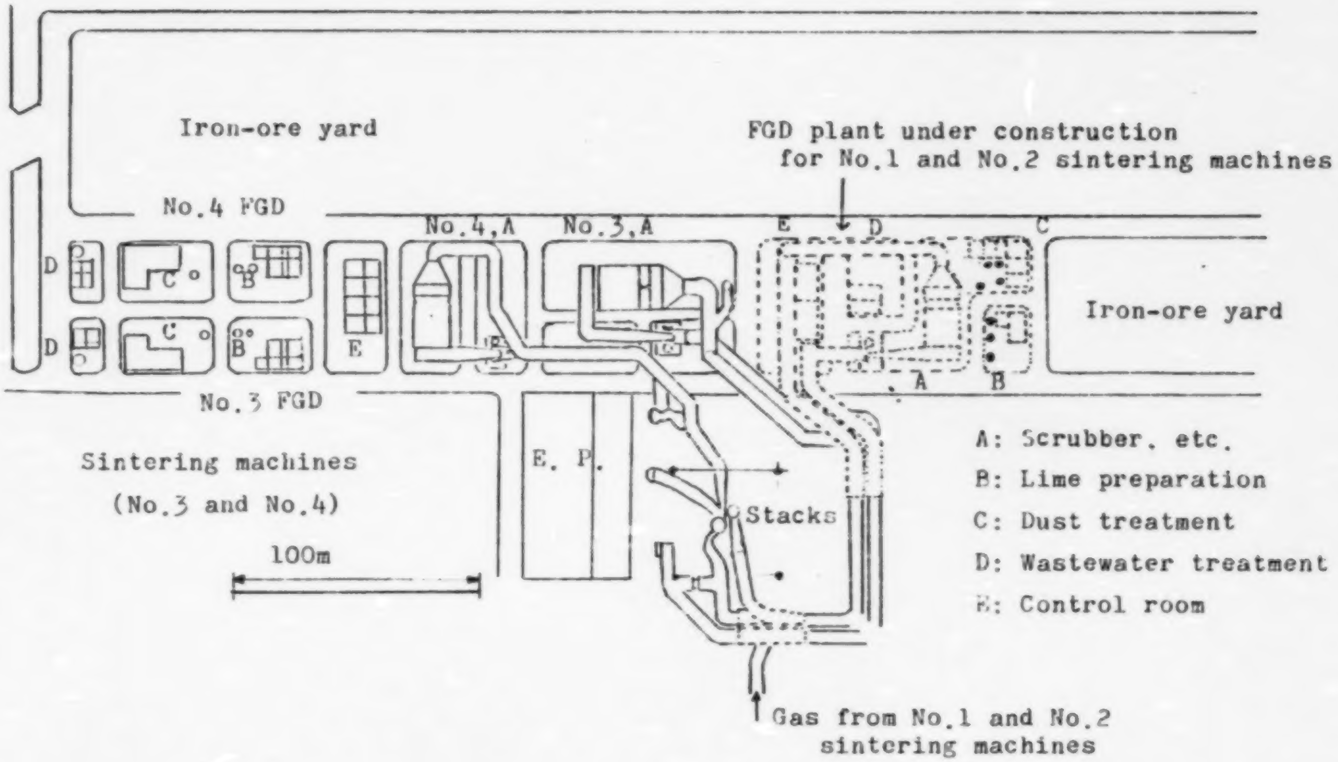

Iron-ore yard
FGD plant under construction
for No.1 and No.2 sintering machines
No.4 FGD
No.4,A
No.3,A
E
D
C
D
D
C
B
C
B
E
Iron-ore yard
No.3 FGD
A
B
Sintering machines
(No.3 and No.4)
E. P.
Stacks
100m
A: Scrubber, etc.
B: Lime preparation
C: Dust treatment
D: Wastewater treatment
E: Control room
Gas from No.1 and No.2
sintering machines

Figure 1 Layout of FGD plants (Mizushima plant, Kawasaki Steel)

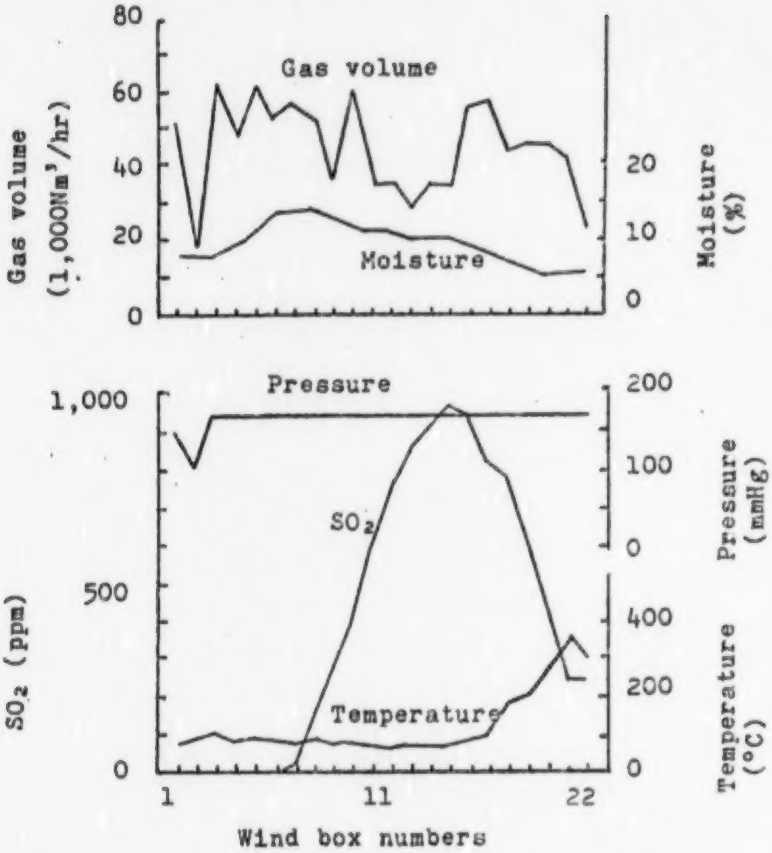


Figure 2 Waste gas from No.4 sintering machine

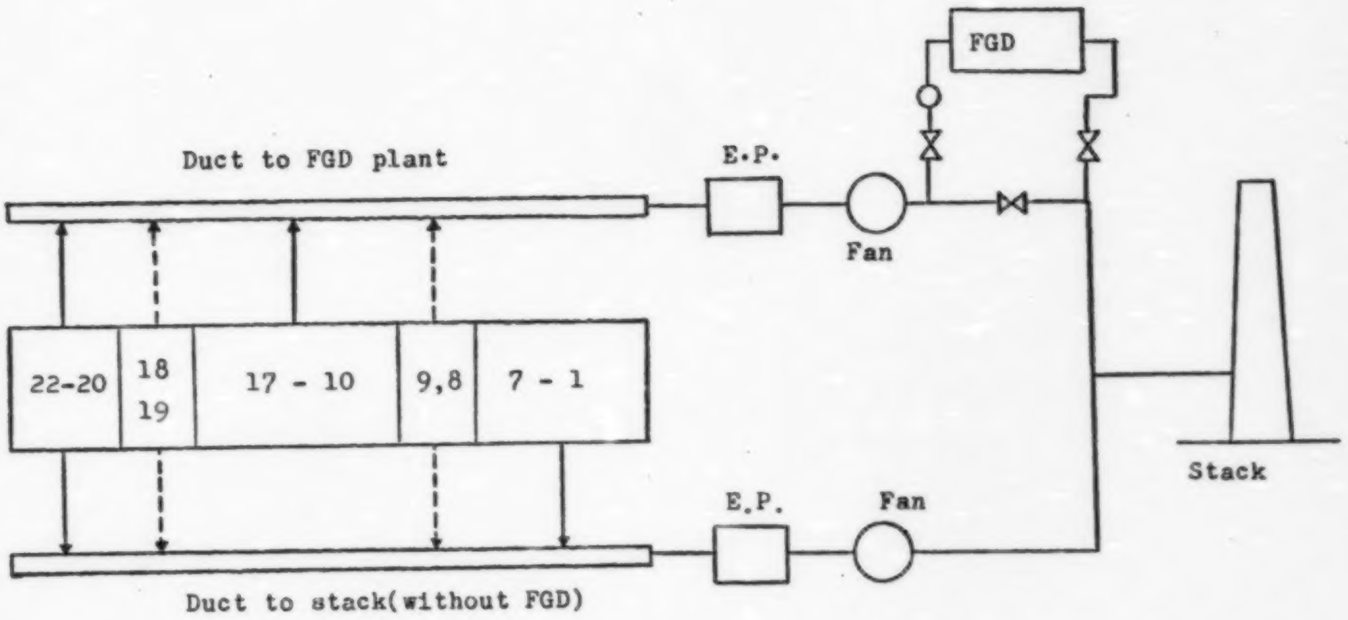

Duct to FGD plant
E.P.
Fan
FGD
22-20
18
19
17 - 10
9,8
7 - 1
Stack
E.P.
Fan
Duct to stack(without FGD)

Figure 3 Gas flow from No.4 sintering machine
(Figures show wind box numbers)

FGD Plant for No. 4 Machine

The plant uses the MHI process with one absorber and has a capacity of treating 750,000Nm^3/hr flue gas (Figure 4). The gas at about 150°C containing 500-1,000ppm SO_2 and about 13.5% O_2 is first cooled to 57°C in a cooler with water sprays and led into a plastic grid packed absorber, where it is treated with a lime slurry at pH 6.4-7.5 at an L/G ratio of 7 liters/Nm^3 (about 50 gallons/1,000scf) and has more than 90% of SO_2 removed. The treated gas passes through a mist eliminator, is heated to about 140°C by after-burning, and sent to a stack. A calcium sulfite slurry discharged from the absorber is acidified to pH 4 by adding sulfuric acid, led into an oxidizer, and oxidized into gypsum by air bubbles generated by a rotary atomizer. The gypsum slurry is sent to a thickener and then centrifuged to less than 10% moisture. The by-product gypsum is sold as a retarder of cement setting. The liquor from the centrifuge is returned to the thickener; the thickener overflow is returned to the absorber after lime is added.

A portion of the circulating liquor of the cooler is neutralized with lime to recover low-grade gypsum. The liquor from the centrifuge is sent to a wastewater treatment system and reused.

The specifications of the FGD unit are shown in Table 3. As the calcium sulfite is oxidized to a considerable extent in the absorber due to the high concentration of oxygen in the gas, the gypsum is recycled to the absorber as crystal seed in order to prevent scaling. To ensure high operability of the plant, spare units have been provided for major pumps and the centrifuge. An automatic system has been installed to shut down and restart the plant with the sintering machine.

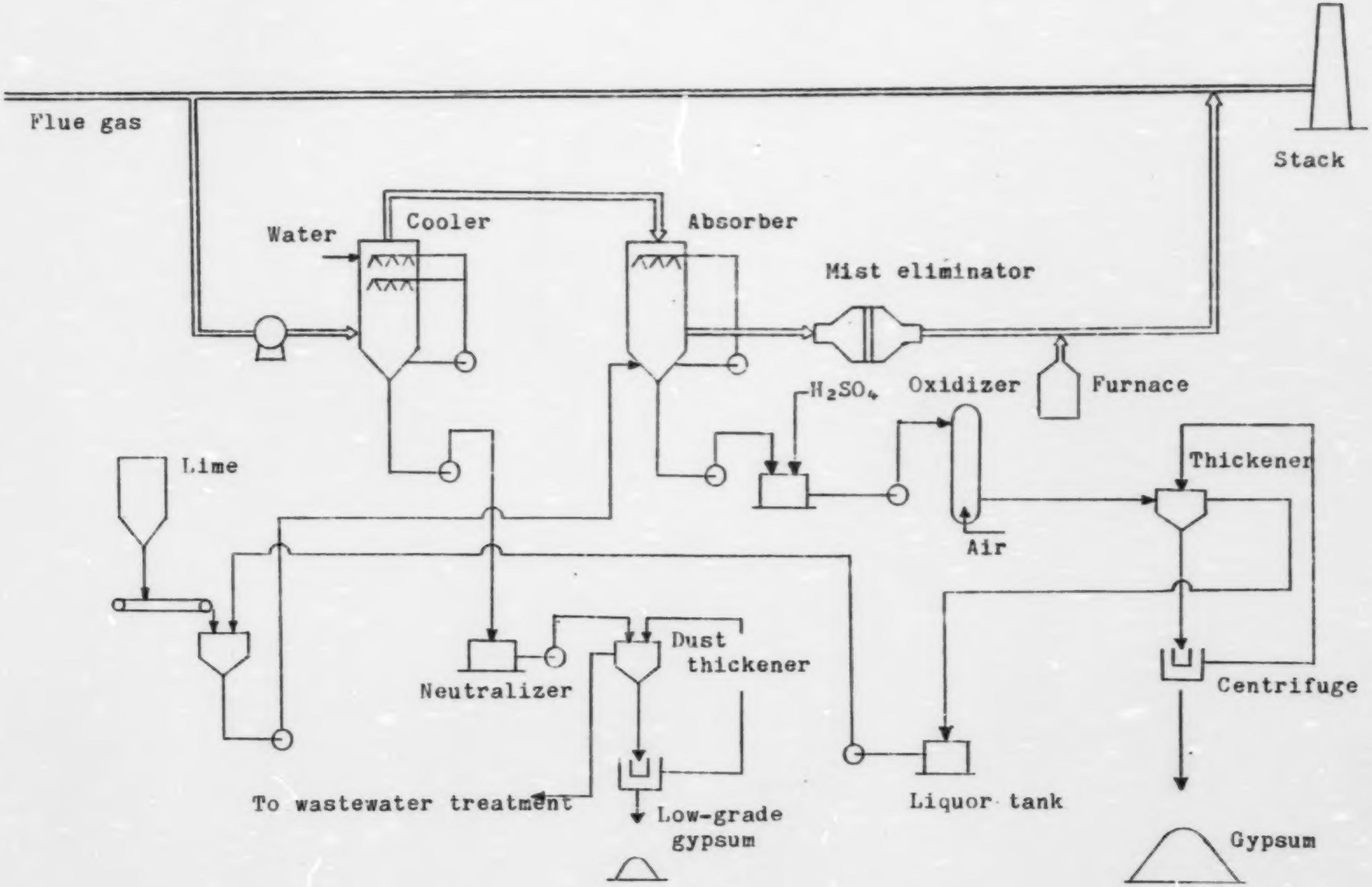

Flue gas
Stack
Water
Cooler
Absorber
Mist eliminator
H_2SO_4
Oxidizer
Furnace
Lime
Thickener
Air
Dust
thickener
Neutralizer
Centrifuge
To wastewater treatment
Low-grade
gypsum
Liquor tank
Gypsum

Figure 4 Flowsheet of MHI lime-gypsum process (one-absorber system)

Table 3. SPECIFICATIONS OF FGD UNIT FOR No. 4 SINTERING MACHINE (Capacity 750,000Nm^3/hr)

Cooler	14m x 8m, 26m high
Absorber	14m x 6.5m, 30m high
Blower	750,000Nm^3/hr, 2,400kW Total pressure 380mmH_2O
Mist eliminator	14m x 13m, 9.5m high
Oxidizer	3.2m diameter, 15.6m high

Performance

The FGD plant for the No. 4 machine went into operation in November 1974. The performance is shown in Figure 5. The gas volume fluctuated from 350,000 to 850,000Nm^3/hr and inlet SO_2 concentration from 400 to 1,100ppm. The SO_2 removal efficiency was 91-98% and the SO_2 concentration at scrubber outlet ranged from 20 to 50ppm. Availability (FGD operation hours per cent of total hours) was about 90% for the first three months because of several minor troubles such as corrosion of the impeller of the cooler circulation pump, stop-up of the lime-slurry pump, and breakage of a fire-brick in the furnace. Those problems were solved and nearly 100% availability was obtained in the next three months. The low availability in May 1975 (about 90%) was due to a shut-down of the sintering machine.

On the average, the gas velocity in the scrubber is about 2.5m/sec and the total pressure drop in the cooler, absorber and mist eliminator is about 120mmH_2O. Lime with less than 1% MgO has been used. The by-product gypsum contains about 7% moisture after being centrifuged and has an average crystal size of about 40 microns. The requirements for the operation are shown in Table 4. At the beginning of the operation, an excessive amount of lime was used to ensure a high SO_2 removal efficiency of over 97%, resulting in the consumption of a considerable amount of sulfuric acid. In later operation,

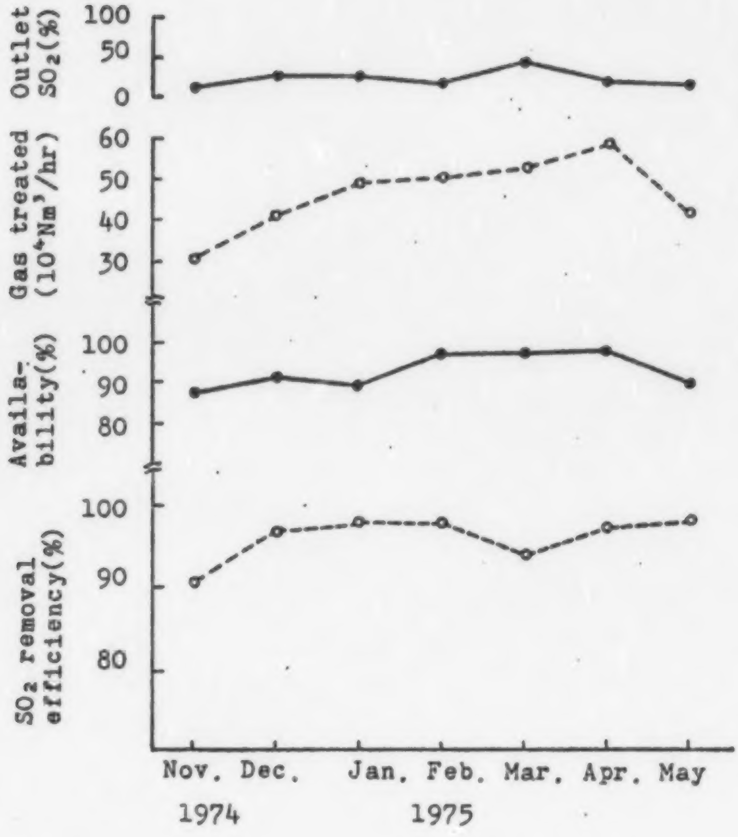


Figure 5 Performance of FGD plant for No.4 sintering machine

slightly less than 1 mole lime to 1 mole inlet SO_2 has been used to obtain about 95% removal and thus the consumption of sulfuric acid has been reduced.

Table 4. REQUIREMENTS OF FGD PLANT FOR No. 4 SINTERING MACHINE

	1974	1975				
	Dec.	Jan.	Feb.	Mar.	Apr.	May
Power (10^3kWhr)	2,000	2,300	2,100	2,500	2,600	2,100
Fuel* (10^6kcal)	7,000	7,000	6,300	7,000	10,000	2,900
Air** (10^3Nm^3)	2,500	1,800	1,900	1,800	1,900	2,300
CaO (t)	500	600	460	530	470	360
H_2SO_4 (t)	24	18	19	13	8	7
Water (10^3m^3)	37	11	24	24	29	19

* For reheating ** For oxidation

3. KASHIMA PLANT, SUMITOMO METAL (MORETANA PROCESS)

Moretana Process

Sumitomo Metal is operating two plants and constructing three more (Table 1), all using the Moretana process developed by Sumitomo jointly with Fujikasui Engineering Co. The process is characterized by the use of the Moretana scrubber fitted with four perforated plates made of stainless steel. The holes range from 6 to 12mm in diameter and the plate thickness from 6 to 20mm. Both dimensions are varied depending on the specific scrubbing conditions. The free space in the cross section ranges from 25 to 50%. The bottom tray serves mainly as a gas distributor and the upper three serve as absorbers. The gas and liquid flows are so adjusted as to maintain a liquor head of 10 to 15mm on each plate. The gas velocity is higher than in usual scrubbers. The design gives extreme turbulence, producing foam layers 400 to 500mm thick, and thus ensures a high SO_2 and dust removal ratio. The mist eliminator is a set of vertical chevron sections mounted in a horizontal duct after the scrubber.

A flowsheet of the process is shown in Figure 6. Gas from a sintering machine is first treated with water in a Moretana scrubber for cooling and to remove more than 90% of dust. Removal of ferric oxide dust is useful in reducing the oxidation in the absorber to ease scale-free operation. The gas is then treated with a limestone slurry (or a mixed slurry of lime and limestone) 10-20% in excess of stoichiometric amount in a second Moretana scrubber to remove more than 95% of SO_2. The limestone contains less than 1% MgO and is ground to pass 325 mesh. The calcium sulfite slurry discharged from the scrubber is sent to a clarifier, and then to a pH adjusting tank where the pH is adjusted to about 4.0 by adding a small amount of H_2SO_4.

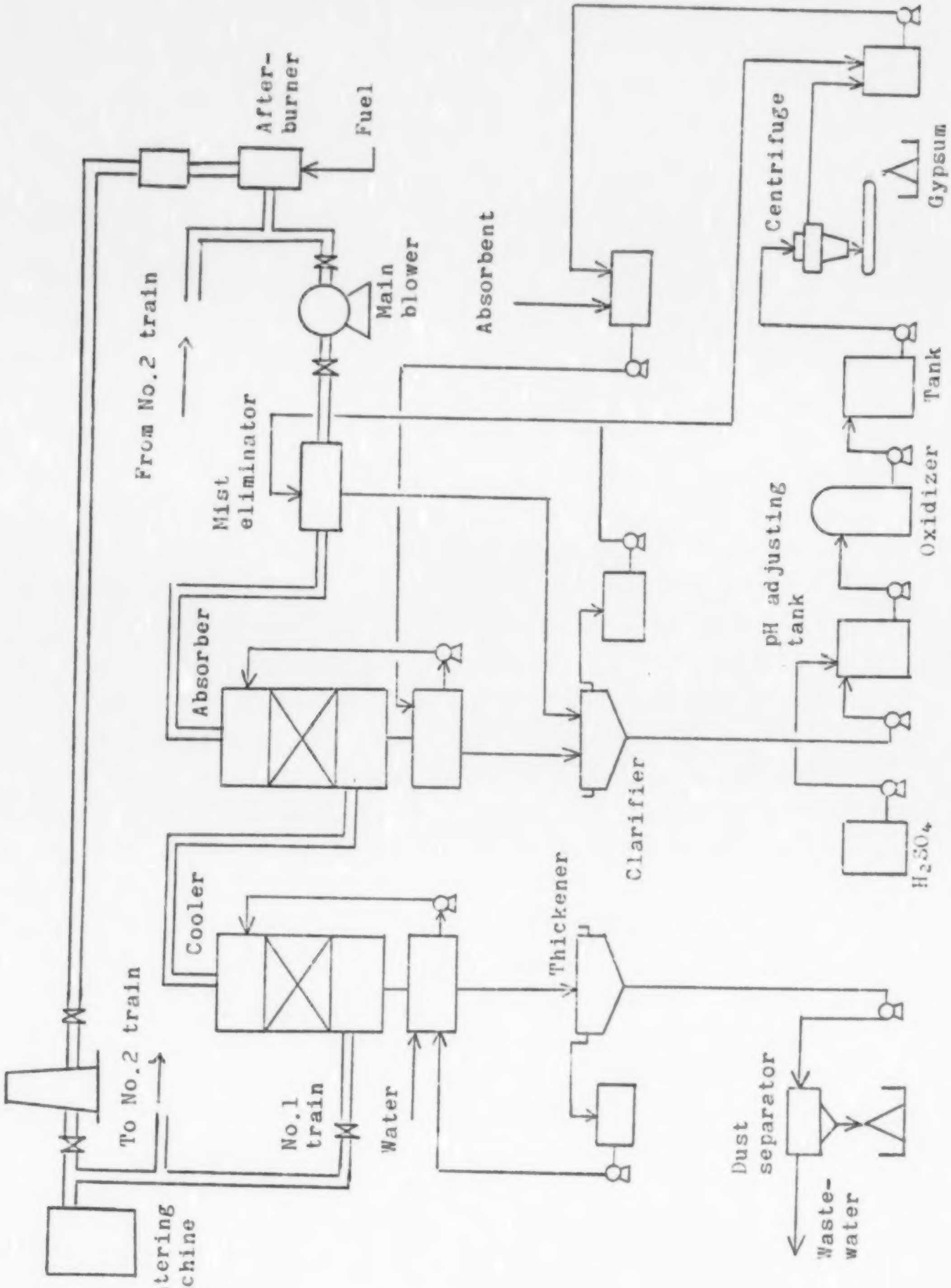

Sintering machine
To No.2 train
No.1 train
Water
Cooler
Thickener
Dust separator
Waste-water
Absorber
Clarifier
H_2SO_4
pH adjusting tank
Oxidizer
Tank
Mist eliminator
From No.2 train
Main blower
After-burner
Fuel
Absorbent
Centrifuge
Gypsum

Figure 6 Flowsheet of Moretana process

The slurry is then sent to an oxidizer developed by Fujikasui to convert calcium sulfite to gypsum. The gypsum slurry is centrifuged, and the filtrate is returned to the absorber.

The discharge from the cooler is sent to a thickener. The overflow is returned to the cooler; the underflow is filtered. The filter cake is returned to the sintering machine and the filtrate is sent to a wastewater treatment system.

Kashima Plant, Sumitomo Metal

The Kashima plant with a capacity of treating 880,000Nm^3/hr gas was started up in September 1975 and has since been in stable operation. All the gas from a sintering machine ranging in flow rate from 350,000 to 880,000Nm^3/hr is treated. The gas contains 200-450ppm SO_2, 14-15% O_2, 6-8% CO_2, 1-1.5% CO, 4-10% H_2O, and 0.15-0.23g/Nm^3 dust at about 150°C. The scrubbing units consist of two trains each with a capacity of treating 440,000Nm^3/hr gas. The Moretana scrubber works normally with a gas velocity between 3 and 5m/sec. When the gas flow rate is low one train only is used. The size of the equipment and operation parameters are shown in Tables 5 and 6.

Table 5. SIZE OF EQUIPMENT

Facility	Number	Size (Specification)
Cooler	2	6.5m (dia.) 24.5m (height)
Absorber	2	6.5m (dia.) 20.5m (height)
Mist eliminator	2	6 x 6m, 2.4m (length)
Oxidizer	2	2.8m (dia.) 5.4m (height)
Centrifuge	4	550kg/hr each

Table 6. OPERATION PARAMETERS

Cooler:	
Space velocity	4.4 to 5.0mm/sec
Circulated liquor	5,000m^3/hr in two towers
Dust content	0.15-0.23g/Nm^3(inlet), 0.008-0.010g/Nm^3(outlet)
Absorber:	
Space velocity	4.4 to 5.0m/sec
Circulated liquor	5,500m^3/hr in two towers
pH of the liquor	6 to 6.5
SO_2 content	250-400ppm (inlet), 3-10ppm (outlet)
Mist eliminator:	
Gas flow	5.5-6.0m/sec
Washing liquor	40m^3/hr in two units
Pressure drop:	
Cooler	220-250mmH_2O
Absorber	120-140mmH_2O
Eliminator	20-25mmH_2O
Total	700-800mmH_2O
By-product gypsum:	
Moisture	6.0-9.5% (after centrifuge)
Crystal size	Larger than 50 microns

Operating hours of the sintering and desulfurization plants are listed in Table 7. Operability of the desulfurization plant has been maintained at 100%.

Table 7. OPERATION OF KASHIMA PLANT

	1975				1976		
	Sept.	Oct.	Nov.	Dec.	Jan.	Feb.	Mar.
Operating hours							
Sintering plant	705	729	683	725	729	654	716
Desulfurization plant	(501)*	729	683	725	729	654	716
Availability of sintering plant	97.9	98.0	94.9	97.5	98.0	93.9	96.3
Operability of desulfurization plant	(100)	100	100	100	100	100	100

* Started operation in September

Performance of the No. 1 train is shown in Figure 7. Inlet and outlet SO_2 concentrations have been 300-450 and 3-15ppm respectively. The L/G ratio in the absorber is about 5 liters/Nm^3 (36 gallons/1,000scf).

Construction and operation costs for the Kashima plant are shown in Table 8.

Table 8. COSTS FOR KASHIMA PLANT

Plant cost ($1,000)	15,000
Running cost ($1,000/month)	
Power	121
Fuel	121
Limestone, chemicals	47
Other	38
Fixed cost ($1,000/month)	286
Total	613
Sintered product (t/month)	283,384
Desulfurization cost ($/t)	2.16

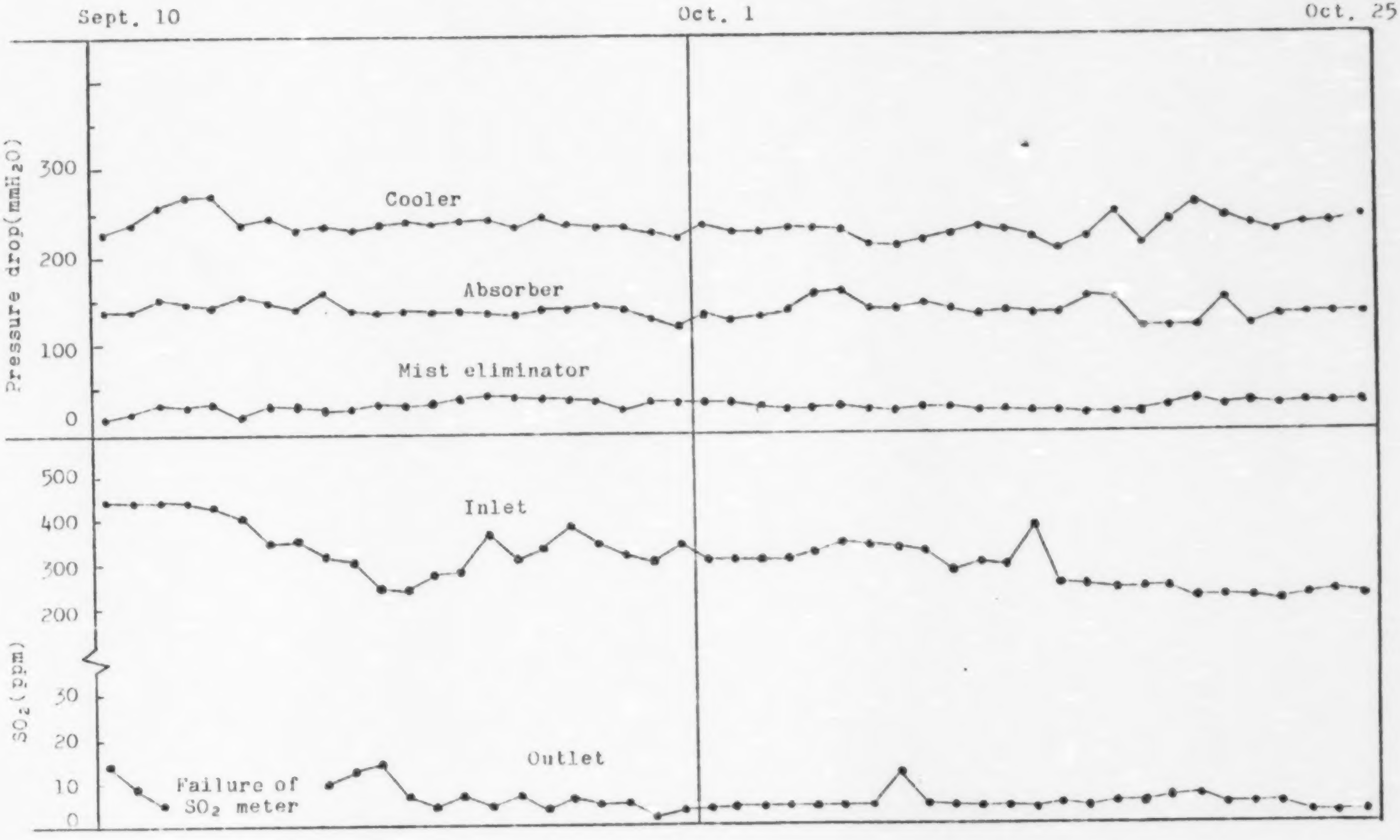


Figure 7 Operation data of No.1 train, Kashima plant

Wakayama Plant, Sumitomo Metal

An FGD plant at Wakayama with a capacity of treating 375,000Nm^3/hr waste gas from a sintering machine started operation in May 1975 and has since been in smooth operation except for a defect in the plastic lining in a cooler which was found at the beginning of the operation and was repaired. Operability of the plant is 98% except for the scheduled shutdown of the sintering plant that normally occurs about every two months. The mist eliminator is washed intermittently (once in 30 minutes) with the circulating liquor and fresh water alternately. The pressure drop in the mist eliminator which is 30mm H_2O at the beginning gradually increases while it is washed with the circulating liquor. When the pressure drop reaches 50mm, fresh water is used in place of the liquor until the pressure drop returns to 30mm. The ratio of liquor to fresh water is about 80 to 20.

4. KOBE STEEL CALCIUM CHLORIDE PROCESS (CAL PROCESS)

Process Description

Kobe Steel has developed a new process using a 30% calcium chloride solution dissolving lime as the absorbent. A pilot plant (50,000Nm^3/hr) has been operated and two commercial plants (Table 1) have just come on-stream to treat waste gas from iron ore sintering plants.

Calcium chloride solution dissolves 6-7 times as much lime as does water. High SO_2 recovery is attained at a low L/G of 3 liters/Nm^3. The flowsheet is shown in Figure 8.

Waste gas is first cooled in a cooler to which a calcium chloride solution (about 5%, from a gypsum centrifuge) is fed to cool the gas to about 70°C and to remove most of the dust. The solution is concentrated here to about 30% and is sent to a scrubber system after dust removal by filtration. The gas is then led into an absorber in which a calcium chloride solution (about 30%, at pH 7 dissolving lime) is sprayed to remove more than 90% SO_2. The gas is then passed through a mist eliminator and sent to a stack. The liquor discharged from the absorber at pH 5.5 containing calcium sulfite is sent through a thickener to a centrifuge to separate most of the solution, which is sent to a tank where calcium hydroxide is dissolved to raise the pH to 7. The calcium sulfite sludge from the centrifuge is repulped with water and some sulfuric acid to produce a slurry at pH 4. The slurry is oxidized by air bubbles into gypsum, which is then centrifuged. The liquor from the centrifuge containing about 5% calcium chloride is returned to the cooler giving no wastewater at all.

Since vapor pressure of the liquor is low, the temperature of the gas after the scrubbing reaches 70°C as compared with the 55-60°C for the usual wet process and thus less energy for reheating is required. The mist eliminator is washed with the circulating liquor. The solubility of gypsum in the liquor is very low (nearly 1/100 of that in water) and the evaporation of the liquor does not cause scaling.

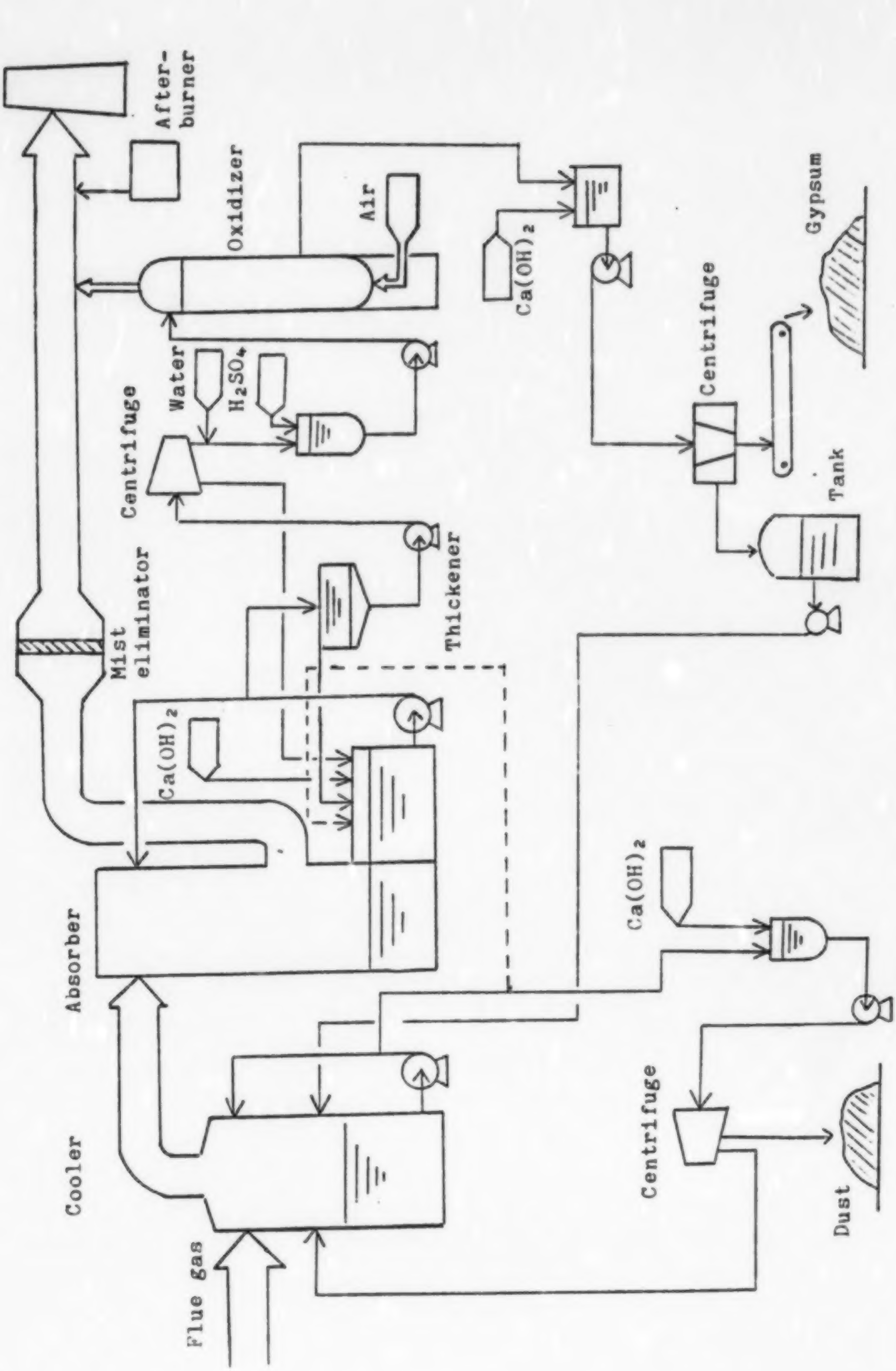

Flue gas
Cooler
Absorber
Mist eliminator
Thickener
Centrifuge
Water
H_2SO_4
Oxidizer
Air
After-burner
$Ca(OH)_2$
$Ca(OH)_2$
$Ca(OH)_2$
Centrifuge
Tank
Gypsum
Centrifuge
Dust

Figure 8 Flowsheet of Cal process

Continuous operation of the pilot plant for about 6 months showed that a soft deposit formed on the wall of the absorber when the L/G ratio was smaller than 1 but the deposit could be washed off by using an L/G larger than 2. A highly corrosion-resistant material is required for the cooler; the lower part where the hot gas comes in is made of titanium.

Amagasaki Plant

The Amagasaki plant has two trains, each with a capacity of treating 175,000Nm^3/hr flue gas at 120°C containing 240-400ppm SO_2, 0.05-0.2g/Nm^3 dust and 14-16% O_2. The plant went into test operation in February 1976. The following problems were encountered during the two months' test run:

Unusual vibration of a centrifuge
Wearing of a control valve
Scaling of pH meter electrode
Breakage of rubber lining in a reducer

Those problems have been solved and the plant went into commercial operation in April 1976. The SO_2 removal efficiency ranges from 91-94%. The dust removal efficiency runs to about 50%. Gas velocity in the absorber is 3m/sec. Total pressure drop in the cooler, absorber and mist eliminator is 190mmH_2O. The L/G ratio is 4.0 in the cooler and 3.0 in the absorber. More than 50% of calcium sulfite is oxidized in the absorber. The by-product gypsum has an average crystal size of about 40 microns and contains about 8% moisture and 0.1% chlorine after being centrifuged. The designed and actual requirements are listed in Table 9.

Table 9. REQUIREMENTS AT AMAGASAKI PLANT

	Designed value	Actual value
Power (kWhr/hr)	1,830	1,372
Water (t/hr)	20	6.2
Steam (t/hr)	2	1.8
Sulfuric acid (kg/hr)	106	50
Slaked lime (kg/hr)	328	150
Calcium chloride (35% solution, kg/hr)	60	50

5. NIPPON STEEL SLAG PROCESS (SSD PROCESS)

Nippon Steel has developed a process which uses converter slag as the absorbent (Figure 9). The slag contains about 40% CaO, 16% SiO_2, 3% MgO, 3% Al_2O_3, and 35% FeO and Fe_2O_3 and has no current uses. Nippon Steel has operated a prototype plant with a capacity of treating 200,000Nm^3/hr waste gas from a sintering plant since 1974. A commercial plant (1,000,000Nm^3/hr) has just started operation.

The process is similar to other lime/limestone-gypsum processes except that it uses no oxidizer. The gas is cooled and led into two absorbers in series to remove 95% of SO_2. The slag is fed to the second absorber to produce a calcium sulfite slurry which is led to the first scrubber and entirely oxidized into gypsum in the scrubber due to a low pH and the presence of much iron compounds which act as a catalyst. The by-product gypsum contains about 40% impurities and has been discarded. The prototype plant has encountered some scaling problem to be solved to ensure a long-term continuous operation. The process may be useful for steel producers who normally have large amounts of useless slag.

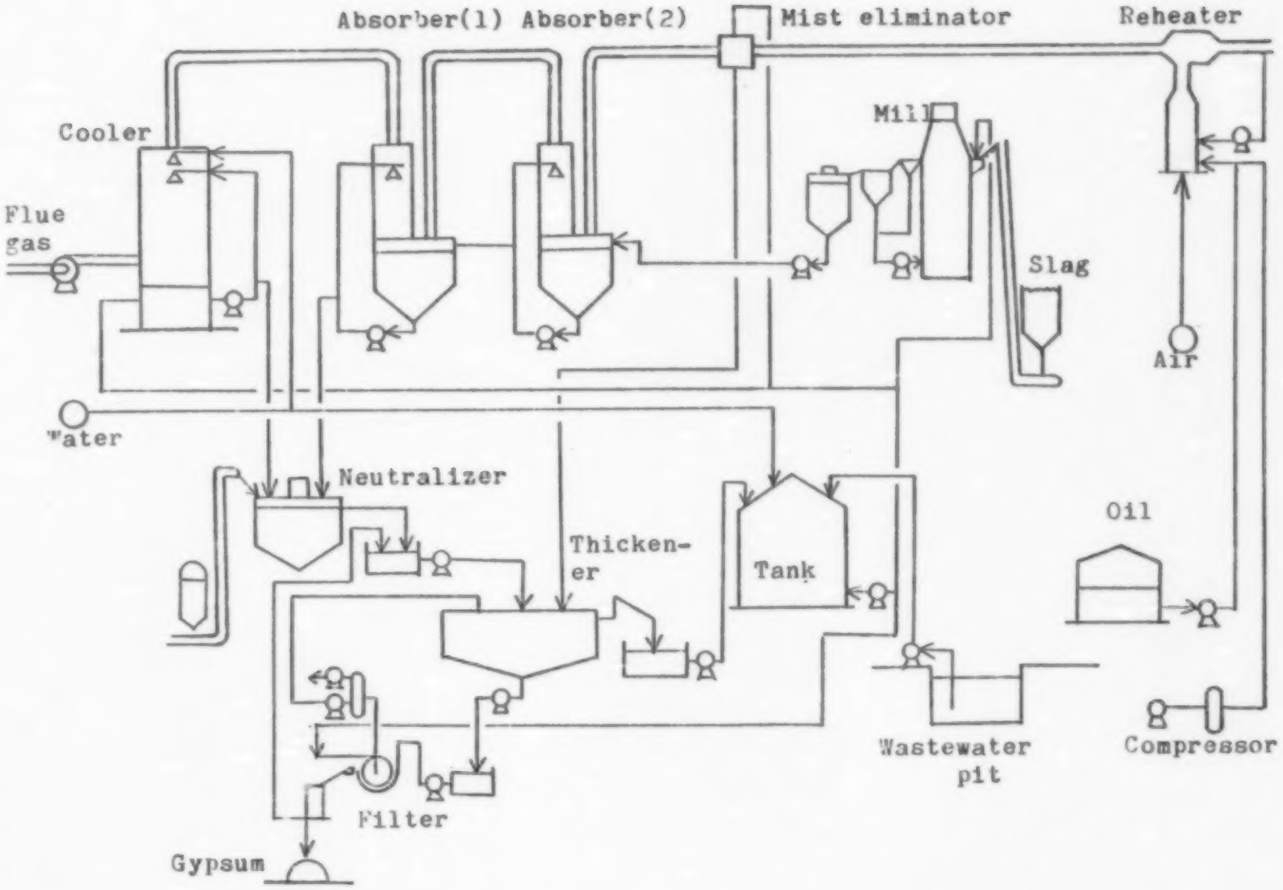

Absorber(1)
Absorber(2)
Mist eliminator
Reheater
Cooler
Mill
Flue gas
Slag
Air
Water
Neutralizer
Oil
Thicken-er
Tank
Wastewater pit
Compressor
Filter
Gypsum

FIGURE 9 - FLOWSHEET OF SSD PROCESS

ADDRESSES OF STEEL PRODUCERS AND PROCESS DEVELOPERS

The addresses of the steel producers and process developers described in this report are listed below:

Kawasaki Steel Co.
1-12-1, Yurakucho, Chiyoda-ku, Tokyo

Sumitomo Metal Co.
5-15, Kitahama, Higashi-ku, Osaka

Kobe Steel Co.
1-36-1, Wakihamacho, Fukiai-ku, Kobe

Nippon Steel Corp.
6-17-2, Ginza, Chuo-ku, Tokyo

Mitsubishi Heavy Industries (MHI)
2-5-1, Marunouchi, Chiyoda-ku, Tokyo

Fujikasui Engineering Co.
1-4-3, Higashigotanda, Shinagawa-ku, Tokyo

APPENDIX C

DESCRIPTION OF RADIAN'S PROCESS SIMULATION MODEL

1.0 INTRODUCTION

Radian has developed a computerized process simulation model for lime/limestone wet scrubbing systems. The program utilizes equipment modules to represent the wet scrubbing system so that different process arrangements can be simulated. The following discussion is a rather brief description of Radian's process simulator as applied to limestone wet scrubbing systems.

2.0 PROCESS DESCRIPTION

A simplified flow diagram of the conceptual limestone scrubbing system is given in Figure 2-1. In this system the incoming flue gas is scrubbed with a limestone slurry in the spray tower. The scrubber bottoms are combined with fresh limestone and recycled slurry from the clarifier in a stirred hold tank where dissolution of limestone and precipitation of calcium sulfate and sulfite occur. The major portion of the tank effluent is returned to the scrubber. The minor portion of the tank effluent is fed to the clarifier, where some additional limestone dissolution and precipitation of calcium sulfate and sulfite may take place. The clarifier underflow is solid waste which exits the process.

The following sections describe the operation of various process components in more detail. Important design relationships are introduced which must be dealt with in some fashion in the process calculation scheme described in Section 3.0.

2.1 Scrubber

The primary purpose of the spray tower is to provide interfacial area for transfer of SO_2 from the flue gas to the alkaline slurry. Gas passing upward through the scrubber is contacted with fine droplets of slurry introduced at the top of the tower via spray nozzles. SO_2 is absorbed by these droplets as they fall through the tower. Limestone present in the slurry droplets may dissolve and the absorbed SO_2 may precipitate as $CaSO_3 \cdot \frac{1}{2}H_2O$, or if it be oxidized, as $CaSO_4 \cdot 2H_2O$.

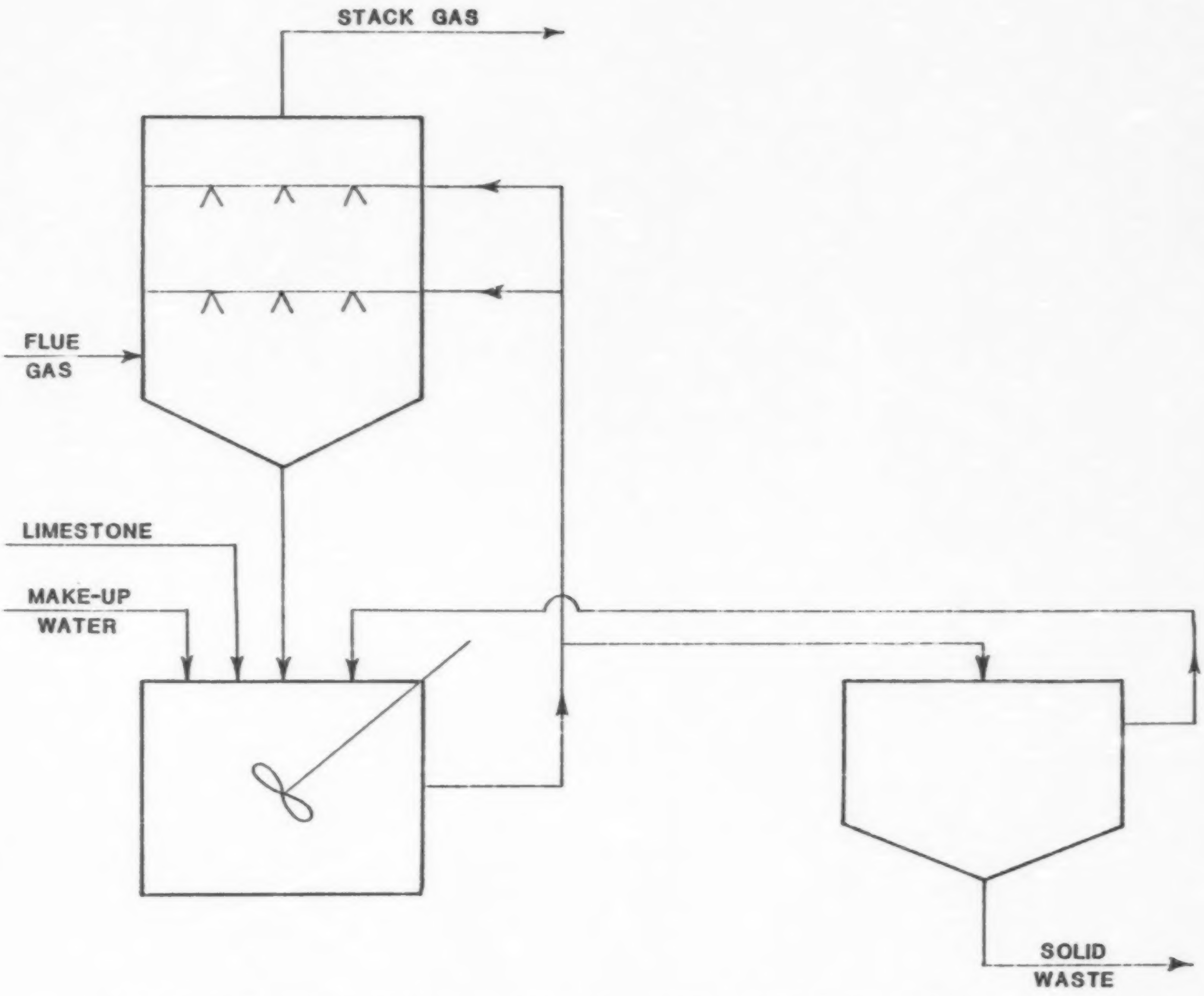

STACK GAS
FLUE GAS
LIMESTONE
MAKE-UP WATER
SOLID WASTE

FIGURE 2-1 FLOW DIAGRAM OF SIMULATED LIMESTONE SCRUBBING PROCESS

A sophisticated mathematical treatment of a spray tower would probably be formulated in terms of mass transfer rate to an individual slurry droplet of a given size. The overall mass transfer performance would then be calculated by considering the number and size distribution of droplets in the total spray. In the absence of such a detailed description of physical phenomena in the spray tower, its overall performance may be discussed using conventional mass transfer terminology derived for packed-tower design. The following discussion is based upon the performance of a spray tower and should not be used to describe other types of gas-liquid contactors.

The usual design problem in application of an absorber is to select a scrubber size and liquid rate to attain a specified absorption efficiency for a given throughput of gas. The design relationship for these parameters may be written (normally for a packed tower) as in Equation 2-1).

$$\frac{K_g aV}{n_G} = \text{N.T.U.} \tag{2-1}$$

Here, K_g (lb mole/hr-ft^2) is the overall gas phase mass transfer coefficient, a (ft^2/ft^3) the interfacial area per unit volume of contactor, n_G (lb mole/hr) the gas flow rate, and N.T.U. the number of overall gas phase mass transfer units. The overall packed height of the tower is calculated as follows:

$$HT = (\text{N.T.U})\ (\text{H.T.U}) \tag{2-2}$$

Here, H.T.U. is the height of an overall gas phase mass transfer unit.

N.T.U. may also be defined in terms of the amount of absorption and the actual and equilibrium mole fractions of the gas being absorbed (SO_2 in this case).

$$\text{N.T.U.} = -\int_{Y_{in}}^{Y_{out}} \frac{dy}{(1-y)(y-y^*)} \qquad (2\text{-}3)$$

When y is very small, as in SO_2 scrubbing where the SO_2 concentration is in the ppm range, then the (1-y) term is approximately equal to 1.

$$\text{N.T.U.} = -\int_{Y_{in}}^{Y_{out}} \frac{dy}{y-y^*} \qquad (2\text{-}4)$$

For systems near atmospheric pressures and with small pressure differentials, the N.T.U. may be expressed in terms of partial pressures.

$$\text{N.T.U.} = -\int_{p_{in}}^{p_{out}} \frac{dp}{p-p^*}$$

$$\approx \frac{p_{in} - p_{out}}{(p-p^*)_{in} - (p-p^*)_{out}} \ln \frac{(p-p^*)_{in}}{(p-p^*)_{out}} \qquad (2\text{-}5)$$

Thus, for a specified SO_2 removal, the required number of transfer units depends on $p^*_{SO_2}$ which is a function of the scrubber liquor composition. If enough available alkalinity is provided in the scrubber slurry, p* will remain small compared to p and sorption will proceed. As the available alkalinity decreases, p* increases and N.T.U. grows large.

Equation 2-1 shows that either K_g, a, or V must be increased to achieve such an increase in N.T.U. A calculated N.T.U. may be used, then as an index of difficulty of achieving specified sorption efficiency for a given set of operating conditions. Calculation of p* and N.T.U. is one of the primary objectives of the present analysis of scrubber performance. The variables which enter into this calculation are the specified SO_2 removal, the amount and composition of scrubber liquor, and the degree of limestone dissolution in the scrubber. The fraction of absorbed SO_2 which is oxidized also affects the equilibrium partial pressure and N.T.U. A more complete design calculation would adjust scrubber design parameters such as gas velocity, height, diameter, to satisfy Equation 2-1. The exact relationship of $K_g a$ to these design variables is not yet known. The results of present calculations may be compared with previous pilot scale data, however, to insure that calculated values of N.T.U. and $K_g aV$ are realistic.

Estimation of mass transfer requirements (N.T.U.) for the scrubber is one goal of an engineering anslysis based on process chemistry. A second important aspect of scrubber performance is the level of supersaturation in the scrubber liquor with respect to $CaSO_3 \cdot \frac{1}{2}H_2O$ and $CaSO_4 \cdot 2H_2O$.

Slurry entering the scrubber will be slightly supersaturated with respect to the solid waste components being precipitated in the hold tank. These are $CaSO_4 \cdot 2H_2O$ and $CaSO_3 \cdot \frac{1}{2}H_2O$. Previous laboratory and pilot scale investigations have shown that a major process problem, scaling, is related to critical levels of supersaturation that should not be exceeded for successful process operation. Under normal operating conditions, the highest level of supersaturation in the process will be reached in the scrubber effluent. Whether or not the critical scaling point is reached will depend on the scrubber liquid-to-gas ratio, the specified SO$_2$ removal, the composition of the inlet slurry and the degree of oxidation and limestone dissolution in the scrubber. Calculated levels of scrubber effluent supersaturation reached under various operating conditions can be compared with experimentally established limits to determine the susceptibility of the scrubber to scaling. Since scaling is primarily a chemical problem and does not depend on assumptions regarding mechanical characteristics of the scrubber, this aspect of the present process analysis will be an important contribution to any preliminary process design. The utility of the simulation model as a design tool and as a method of predicting scaling conditions was demonstrated at a 3 MW limestone scrubbing pilot plant located at Pennsylvania Power and Light's Sunbury Station. The simulation model was used to set process operating parameters. To verify the model's accuracy, the system was operated at conditions in which the model predicted scaling to occur. At those conditions the system scaled. The reliability of the model was shown by the fact that the pilot plant operated throughout its life in an unscaled mode.

2.2 Hold Tank

The function of the process hold tank is to provide adequate reaction time in a well mixed environment for sufficient dissolution of limestone and precipitation of calcium sulfate and sulfite to occur. Since the hold tank is a well mixed vessel, the composition of the output stream will be essentially the same as the composition of the material in the tank.

The performance of a hold tank may be calculated by simultaneous solution of the chemical reaction and mass transfer rate equations describing the phenomena taking place in the vessel. Laboratory and pilot scale investigations have provided quantitative descriptions of precipitation and dissolution reactions important to a limestone scrubbing process.

The rate expression which is used for the hold tank is given in Equation 2-4.

$$R = k\, A\, n\, V\, K_{sp}\, (RS - 1) \qquad (2\text{-}6)$$

Here, R is the precipitation rate, k is the rate constant, A is a proportionality constant, n is the flow rate of the solid into the tank, V is the hold tank volume, K_{sp} is the solubility product constant, and RS is the relative saturation. The relative saturation is the product of the activities of the species which react to produce the precipitating solid divided by the solubility product constant, as shown in Equations 2-5 and 2-6 for calcium sulfate.

$$Ca^{++} + SO_4^{=} + 2H_2O \rightleftarrows CaSO_4 \cdot 2H_2O \downarrow \qquad (2\text{-}7)$$

$$RS = \left[a_{Ca^{++}} \cdot a_{SO_4^{=}} \cdot a^2_{H_2O(\ell)}\right] / K_{spCaSO_4 \cdot 2H_2O} \qquad (2\text{-}8)$$

For precipitation to occur, the relative saturation is greater than one, and the rate R is positive. For dissolution to occur the relative saturation is less than one and the Rate R is negative. At equilibrium, the relative saturation is equal to one, and the rate is zero.

Several factors influence these solid-liquid mass transfer rates. One factor of major importance is the hold tank volume. If the volume is increased for a given relative saturation and flow rate, the precipitation and dissolution rates will increase. If the volume is increased and the rates are to remain the same, the relative saturations must move closer to one, indicating a closer approach to equilibrium.

Another factor which will affect the rates is the flow rate of the solid into the tank (i.e., slurry concentration). This flow rate controls the area on which precipitation and dissolution may occur. Increasing the area for a given relative saturation will increase the rate.

Variation of the proportionality constant A and the rate constant k is known to occur with the particle size of precipitating or dissolving solids. The exact functionality of this change is unknown. For purposes of investigating the effect of such changes, the rate constant k can be changed in process calculations.

The principle criteria for hold tank design in limestone scrubbing processes is that the hold tank effluent supersaturation be low enough to prevent scaling conditions from developing as the slurry is recycled through the scrubber. A secondary requirement is that the tank be large enough to dissolve limestone for reasonable concentrations of limestone in the slurry. A larger hold tank will improve limestone utilization.

The design and function of the hold tank interact with the scrubber operating parameters. Up to a point, increasing the hold tank size will decrease the required scrubber liquor rate at least from the standpoint of supplying sufficient alkalinity and keeping the scrubber effluent supersaturation at a safe level.

2.3 Clarifier

The clarifier is intended primarily as a solid-liquid separation device; however, some additional solid-liquid mass transfer may take place here. One reaction of possible significance that could occur in the clarifier is sulfite oxidation due to the large surface area available for transfer of oxygen from the atmosphere. This oxidation could affect the levels of sulfite and sulfate supersaturation in the scrubbing system.

The clarifier removes waste solids and some liquid from the process. A major process variable associated with the clarifier is the amount of liquid leaving the process with the waste solids. Soluble species such as chloride, sodium, and magnesium are introduced to the system as trace components in the flue gas, fly ash, and limestone. Since the liquid carried with waste solids is the only route by which soluble species can leave the system, the concentration of these soluble salts in the process liquor will be inversely proportional to the liquor content of the waste solids. These soluble species cause significant changes in process chemistry through their interaction with ions that actually participate in the absorption, dissolution, and precipitation steps.

3.0 PROCESS MODEL

This section describes the computational scheme or model used to estimate steady-state operating conditions for a limestone scrubbing process.

The important physical phenomena and process rate steps introduced in Section 2 are formulated mathematically. Certain simplifying assumptions are made which lead to practical yet meaningful solutions describing the performance of the process.

The Radian process model is a gorup of computer programs for simulating aqueous inorganic chemical processes. The foundation of the model is the ability to predict vapor-liquid-solid mass transfer rates and chemical equilibrium for the $CaO\text{-}MgO\text{-}Na_2O\text{-}SO_2\text{-}CO_2\text{-}SO_3\text{-}N_2O_5\text{-}HCl\text{-}H_2O$ system. The process model performs unit operation calculations and other engineering manipulations based upon (1) rate and equilibrium calculations and (2) process and equipment data which define the process flow scheme and characterize each of the individual process units.

The programs which make up the process model may be grouped into five major subdivisions: (1) rate and equilibrium calculation programs, (2) equipment subroutines which model each process unit and process input stream, (3) an executive system which interconnects the equipment subroutines to form an analog of the process flow diagram and controls the sequence of computer operations, (4) convergence routines which force convergence of the model iterative parameters, and (5) print routines which print out stream and process data.

3.1 Process Model Flow Sheet

The main flow scheme used in this group of simulations is given in the symbology of the Radian process model as Figure 3-1. Subroutines FLUGS1, ALKINP, and WTRMKP simulate process inputs of flue gas, limestone, and make-up water, respectively. Subroutine SCRUB5 models the spray tower and RATHD1 simulates a stirred holding tank. Subroutines DIVDER and DIVDR2 simulate stream splitters, and FILTER models a clarifier-filter system. Subroutine SYSTB1 is an ancillary routine which performs material balance calculations around the entire system.

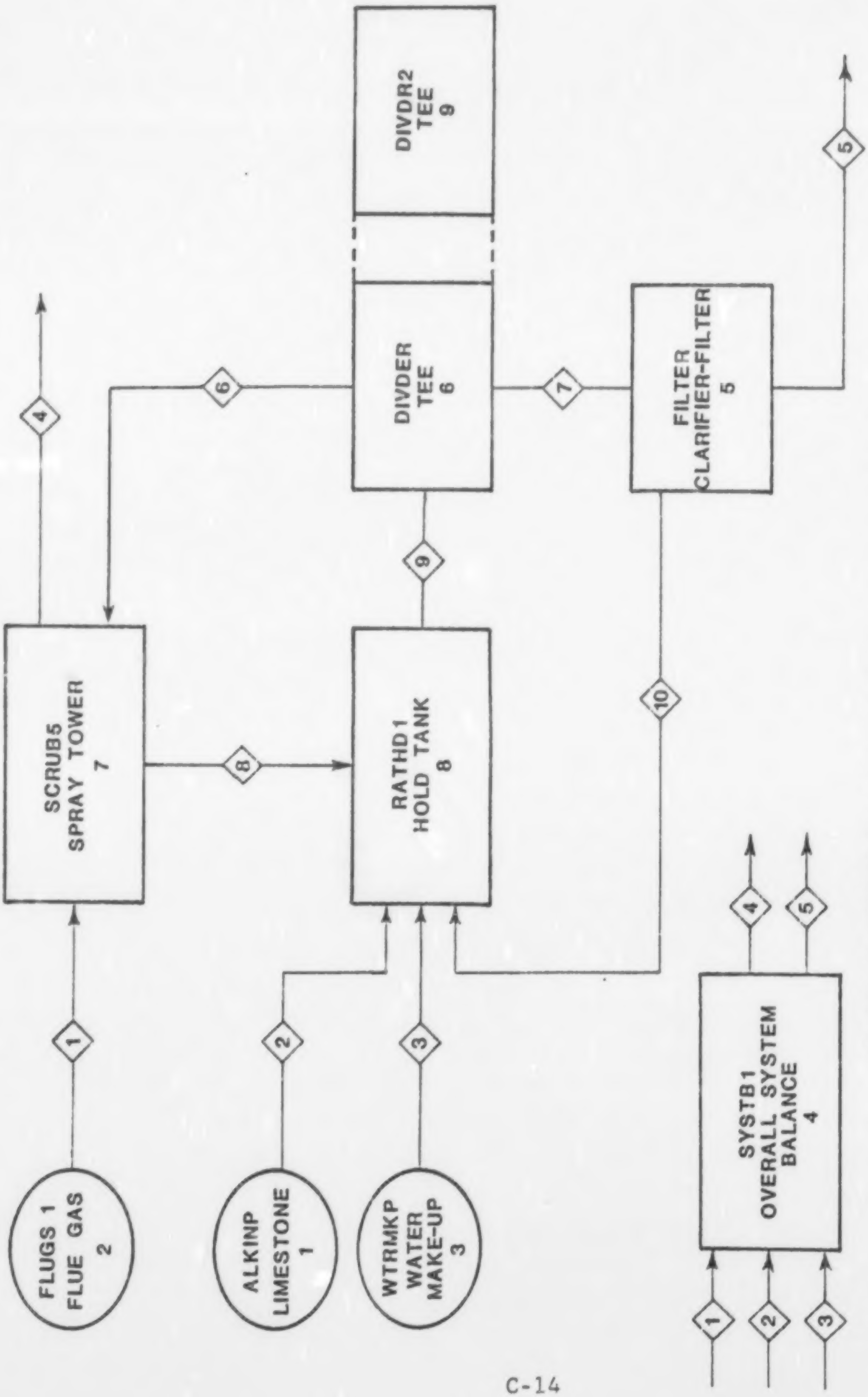
Several assumptions have been incorporated into this simulation system.

(1) The spray tower behaves as an adiabatic countercurrent contacting device and the equilibrium partial pressure of SO_2 above the scrubber feed is negligible.

(2) The partial pressures of CO_2 and H_2O in the gas leaving the scrubber are in equilibrium with the scrubber liquid at the scrubber temperature.

(3) No precipitation occurs in the scrubber.

(4) The temperature of the recirculating process liquor streams is fixed at the adiabatic saturation temperature of the scrubber effluent liquor.


FLUGS 1
FLUE GAS
2
ALKINP
LIMESTONE
1
WTRMKP
WATER
MAKE-UP
3
SCRUB5
SPRAY TOWER
7
RATHD1
HOLD TANK
8
DIVDER
TEE
6
DIVDR2
TEE
9
FILTER
CLARIFIER-FILTER
5
SYSTB1
OVERALL SYSTEM
BALANCE
4
1
2
3
4
5
6
7
8
9
10

ORDER OF PROCESS CALCULATIONS: 1, 2, 3, 4, 5, 6, (7, 8, 9) *

FIGURE 3-1 CONCEPTUAL DESIGN FLOW PLAN

(5) The temperature of the water make-up has no effect upon the circulating liquor temperature.

(6) Ionic reactions taking place in the liquid phase are rapid and thus are at equilibrium.

(7) The holding time of the clarifier (modeled by the FILTER subroutine) is sufficiently long that solid-liquid equilibrium is achieved.

(8) The hold tank closely approaches an idealized backmixed vessel.

(9) Neither chemical change nor phase separation occurs to a process stream except in a process unit.

The two assumptions involving the temperature of the process liquor streams should be good approximations. Previous pilot unit work indicates that the scrubber effluent liquor streams closely approach the adiabatic saturation temperature. In simulation cases previously conducted, the water make-up stream was on the order of 0.5 percent of the scrubber feed stream and thus would have a negligible effect upon changing the process liquor temperature. Heat loss to the surroundings should be small in most instances. Assumptions one and three are good approximations for a short-residence-time contactor with a high liquid-to-gas ratio such as a spray tower. Assumption seven listed above is probably not a good assumption in that it does not reflect the true situation in a clarifier. This modeling assumption may be justified from the standpoint that solid-liquid equilibrium represents the maximum chemical change that can occur across the clarifier. Thus, the actual composition of the clarifier effluent streams will lie between the composition of the clarifier feed and the effluent composition as determined by solid-liquid equilibrium. In general, this chemical change will be small.

3.2 Equipment Subroutines

This section gives a brief description of the major routines used in the process simulations. The routines may be grouped into four sections: (1) input routines (FLUGS1, ALKINP, and WTRMKP), (2) unit operation routines (SCRUB5, RATHD1), (3) material balance routines (SYSTB1), and (4) minor process unit routines (DIVDER, DIVDR2, and FILTER).

FLUGS1 - This routine simulates a flue gas input stream with entrained fly ash. It reads data cards for the temperature, pressure, and flow rate of the gas stream and the weight rate of the fly ash, all in English units , as well as the mole fraction composition of the gas and the weight fraction composition of the solid. The calculations performed are to convert these data to program units and assign it to the specified output stream.

ALKINP - This routine simulates an alkali input stream, which was limestone for all cases run. It reads data cards for the weight rate and weight fraction composition of the limestone. The calculations performed are to convert these data to program units and assign it to the specified output stream.

WTRMKP - This routine simulates a water input stream. It reads data cards for the weight composition of the water stream and converts these data to molality in the specified output stream. This routine does not assign the flow rate of the output stream.

SCRUB5 - This routine simulates a vapor-liquid contactor. It is assumed that (1) the scrubber is adiabatic, (2) CO_2 and H_2O are in vapor-liquid equilibrium, and (3) no precipitation occurs. The flue gas flow rate, composition, and amount of SO_2 removal required is input to this routine as is the composition and flow rate of the scrubber slurry feed stream. Since quantitative prediction of the amount of limestone that will dissolve in a spray tower is not yet feasible, a fraction of incoming limestone (and fly ash, if desired) is assumed to dissolve. The type of contactor (co-current, countercurrent, and back-mixed liquor) may also be specified. The mass transfer coefficients reported here are for countercurrent spray tower operation.

The routine calculates the number of transfer units (and thus K_GaV) required to achieve the specified SO_2 removal using the specified feed rate and composition. The composition of the outlet slurry is also calculated. As noted in Section 2.0, the supersaturation of this stream is of major interest.

RATHD1 - This routine simulates a well-mixed holding tank. All input streams must be completely known along with the tank volume and the rate constants for the solid-liquid mass transfer of limestone, $CaSO_4 \cdot 2H_2O$, and $CaSO_3 \cdot \frac{1}{2}H_2O$. The output stream is calculated by simulataneous solution of rate, material balance, and equilibrium relationships.

Hold tank performance can be modified by adjusting the tank volume, the slurry solids concentration, or by changing the precipitation and dissolution rate constants.

SYSTB1 - This routine performs a mass and energy balance about the process based on complete knowledge of the flue gas and limestone streams, the composition of the make-up water stream, and the following system parameters: (1) the fraction of the SO_2 in the flue gas which is absorbed, (2) the fraction of the absorbed SO_2 which is oxidized to SO_3, (3) the pressure drop across the scrubber, (4) the particulate removal efficiency of the scrubber, and (5) the desired filter bottoms pH. Additional information is taken from other routines as need be.

Based on total removal of SO_3 and HCl, the specified fractional removal of SO_2, and a guess at the loss of CO_2, the non-aqueous portion of the scrubber exit gas is calculated. An adiabatic heat and material balance is then performed to find the water in the scrubber exit gas, assuming that the liquid circulating through the scrubber and the exit gas are at the adiabatic saturation temperature.

Based on (1) the weight fraction solids in the filter bottoms; (2) complete information for the flue gas, scrubber exit gas, and limestone streams; and (3) the composition of the make-up water streams, the filter bottoms flow rate and composition, and the make-up water flow rate can be calculated.

DIVDER - This routine simulates a stream splitter. Based on the flow rate for the first output stream and complete information about the second output stream, complete information for the input stream and the first output stream are calculated.

DIVDR2 - This routine also simulates a stream splitter. The difference from DIVDER is that in DIVDR2 the feed stream is completely known and the flow rate of the first output stream is known. Based on these data, complete information for both output streams is calculated.

FILTER - This routine models a solid-liquid separator. Input data are the weight fraction solids in the feed and bottoms streams and the separation efficiency, i.e., the fraction of the feed solids which are transferred to the bottoms.

3.3 Calculation Sequence

In the execution of a simulation computer run, the executive system makes three passes through each of the equipment subroutines which is used in the simulation. In the first pass the data cards are read, the input data are printed, and any necessary initialization (such as assigning the weight fraction solids in the filter bottoms in the FILTER routine) is performed. In the second pass the simulation calculations are performed as follows.

(1) The input routines FLUGS1, ALKINP, and WTRMKP are called. These routines perform all of their operations on the first pass and do not play an active part on the second pass.

(2) Subroutine SYSTB1 calculates the scrubber exit gas and the filter bottoms stream.

(3) Subroutine FILTER calculates the filter feed and filter overhead.

(4) Subroutine DIVDER calculates a first guess at the scrubber slurry feed stream and the hold tank effluent as having equilibrium compositions.

(5) Subroutine SCRUB5 calculates the scrubber bottoms stream.

(6) Subroutine RATHD1 calculates the tank effluent based on rate calculations.

(7) Subroutine DIVDR2 recalculates the scrubber feed and filter feed streams based on the hold tank effluent.

(8) Steps 5, 6, and 7 (subroutines SCRUB5, RATHD1, and DIVDR2) are iterated until the compositions and flow rates of the streams involved approach their steady-state values.

This completes the second pass. At this point, the executive system prints complete stream data for all process streams. After the stream print, a final pass is made through the equipment subroutines, and any ancillary output which was not printed in the stream print, such as $K_g aV$ and N.T.U. in the scrubber is printed.

4.0 PROCESS SIMULATION CASE

The following pages represent a typical process simulation of a limestone wet scrubbing system. The computer output for the conceptual design of the limestone scrubbing system on the sinter plant with 39 percent windbox gas recycle was chosen. The stream values shown were those used to prepare the material balance which was presented in Section 5.1 of this report. The flow plan used for the simulation case was previously given in Figure 3-1 of this Appendix.

RADIAN
CORPORATION

```
22 JUN 76    11124157.151                                    TEMPERATURE    57.075 DEG. C.
                                 INPUT SPECIES (MOLES)

                   H2O  = 9.78174+04              HCL  = 4.20609+00
                   CAO  = 1.57154+03              CO2  = 2.81233+02
                   MGO  = 1.52581+02              N2O3 = 0.00000
                   NA2O = 1.52833+00              N2O5 = 3.07505-02
                                                  SO2  = 3.81911+02
                                                  SO3  = 9.13598+02

                              AQUEOUS SOLUTION EQUILIBRIA

             COMPONENT          MOLALITY          ACTIVITY      ACTIVITY COEFFICIENT

              H+                5.412-07          4.574-07            8.451-01
              H2O                                                     9.994-01
              H2CO3             1.338-03          1.350-03            1.009+00
              HCO3-             1.849-03          1.499-03            8.109-01
              HNO3              1.028-12          1.037-12            1.009+00
              H2SO3             1.314-08          1.325-08            1.009+00
              HSO3-             2.458-04          1.993-04            8.109-01
              HSO4-             6.372-07          5.165-07            8.105-01

              CA++              1.164-02          5.163-03            4.434-01
              CAOH+             3.325-08          2.695-08            8.105-01
              CAHCO3+           2.178-04          1.765-04            8.105-01
              CACO3             2.711-06          2.735-06            1.009+00
              CANO3+            9.054-07          7.338-07            8.105-01
              CASO3             3.149-04          3.178-04            1.009+00
              CASO4             7.055-03          7.119-03            1.009+00

              MG++              6.154-04          2.555-04            4.036-01
              MGOH+             3.568-08          2.879-08            8.105-01
              MGHCO3+           5.729-05          4.641-05            8.105-01
              MGCO3             2.413-07          2.434-07            1.009+00
              MGSO3             6.623-05          6.661-05            1.009+00
              MGSO4             2.642-04          2.665-04            1.009+00

              NA+               1.829-03          1.489-03            8.207-01
              NAOH              7.178-11          7.244-11            1.009+00
              NAHCO3            2.193-05          2.213-05            1.009+00
              NACO3-            4.335-08          3.514-08            8.105-01



              NANO3             1.501-09          1.515-08            1.009+00
              NASO4-            5.082-05          4.117-05            8.105-01

              OH-.              2.216-07          1.796-07            8.105-01
              CL-               2.389-03          2.322-03            9.862-01
              CO3--             5.401-07          2.338-07            4.324-01
              NO3-              3.508-05          2.789-05            7.739-01
              SO3--             3.897-05          1.685-05            4.324-01
              SO4--             1.093-02          4.364-03            4.284-01

             COMPONENT          MOLALITY       ACTIVITY PRODUCT  RELATIVE SATURATION

              CA(OH)2(S)        0.000             1.665-16            5.981-11
              CACO3(S)          1.008-01          1.206-09            8.602-01
              CASO3(S)          2.224-01          8.597-08            1.361+00
              CASO4(S)          5.152-01          2.251-05            1.268+00
              MG(OH)2(S)        0.000             9.748-18            1.852-06
              MGCO3(S)          0.000             6.655-11            6.337-06
              MGSO3(S)          0.000             4.922-09            1.162-04

             PCO2 = 8.45207-02  ATM.
             PSO2 = 3.34731-09  ATM.
             PHNO3 = 1.75722-17  ATM.

                             MOLECULAR WATER = 1.71264+03 KGS.

      PH =  6.3397           IONIC STRENGTH = 4.07796-02        RES. E.N. =  -1.799-08

                                     RATE PARAMETERS

            COMPONENT          INPUT RATE         RATE CONSTANT         PRECIPITATION
                               GMOLE/SEC          KG H2O/SEC          RATE, GMOLE/SEC

             CACO3(S)          2.759+02            6.801+01              -1.543+02
             CASO3(S)          3.795+02            1.293+01               1.036+00
             CASO4(S)          6.805+02            3.072+01               1.798+00
            SINTER PLANT ( 39% RECYCLE ) LIMESTONE SCRUBBING SYSTEM

                            PROCESS DESCRIPTION

 EQUIP. NO.      EQUIP. NAME     INPUT STREAMS          OUTPUT STREAMS

     1             ALKINP                                 1
     2             FLUGS1                                 2
     3             WTRMKP                                 3
     4             SYST01         2   3   1               4   5
     5             FILTER         7                      10   5
     6             DIVDER         9                       6   7
     7             SCRUB5         2   6                   4   8
     8             RATHO1         1   8  10   3           9
     9             DIVDR2         9                       6   7

                        ORDER OF PROCESS CALCULATIONS

        1,2,3,4,5,6(7,8,9)*

        RECYCLE LOOP FROM   7  TO   9


SYSTEM AND EQUIPMENT PARAMETERS

    SYST01, EQUIPMENT NUMBER  4
         SO2 ABSORBED =   90.40 %        NO  ABSORBED =      .00 %
         SO2 OXIDIZED =   70.00 %        NO2 ABSORBED =      .00 %
         LIME SOLIDS HYDRATING IN SYSTEM*  CAO =  100.00 %   MGO =  100.00 %
         PRESSURE DROP CONSTANTS*  GAMMA1 =  0.000      GAMMA 2 =  0.000      GAMMA 3 =  0.000
         PARTICULATE REMOVAL CONSTANTS*  BETA 1 =  1.204+00  BETA 2 =  0.000      BETA 3 =  0.000
         INITIAL VALUE OF XA(CO2) =  0.0000
         SOLID WASTE PH LIMITED TO  6.00 & 0.5

    FILTER, EQUIPMENT NUMBER  5
         FEED WT FR SOLIDS =  12.00 %   UNDERFLOW WT FR SOLIDS =  40.00 %
         SEPARATION EFFICIENCY = 100.00 %

    DIVDER, EQUIPMENT NUMBER  6
         STREAM NUMBER  5 FLOW =  2.7451+04 GAL/MIN

    SCRUB5, EQUIPMENT NUMBER  7
         BACKMIXED OPERATION       PRECIPITATION NOT ALLOWED
         LIME SOLIDS HYDRATING IN SCRUBBER*  CAO = 100.00 %   MGO = 100.00 %
         FRACTIONS OF SOLIDS AVAILABLE FOR REACTION*
                        CA(OH)2   MG(OH)2     CACO3      MGCO3       MGSO4       NA2O        NACL      LIMESTONE
              SLURRY*      .00 %      .00 %      .75 %      .00 %       .00 %       .00 %       .00 %       .00 %
              FLUE GAS*  100.00 %  100.00 %      .00 %      .00 %       .00 %   100.00 %   100.00 %       .00 %

    RATHO1 EQUIPMENT NUMBER  8
         VOLUME =  5.0000+04 CU FT   =  3.7403+05 GAL
         RATE CONSTANTS (GMOLE/SEC CM SQ)*
              CACO3 =  7.5000-02  CASO4 =  2.0000-05  CASO3 =  1.5300-04  LIMESTONE =  0.0000
                                    CRYSTAL AREA SPECIFIED
         CACO3 = 4.6500+03     CASO3*1/2H2O = 7.5000+03     CASO4*2H2O = 4.0000+03

    DIVDR2, EQUIPMENT NUMBER  9
         STREAM NUMBER  5 FLOW =  2.7461+04 GAL/MIN


INPUT STREAM DATA

    ALKINP, EQUIPMENT NUMBER  1
         FLOW     6.3340+01 LB/MIN             CASO3    0.0000     WT %
         CAO      0.0000     WT %              MGSO3    0.0000     WT %
         MGO      0.0000     WT %              CASO4    0.0000     WT %
         CACO3    9.7000+01  WT %              MGSO4    0.0000     WT %
         MGCO3    0.0000     WT %              INERT    3.0000+00  WT %
         CA(OH)2  0.0000     WT %              MG(OH)2  0.0000     WT %

    FLUGS1, EQUIPMENT NUMBER  2*  135. DEG F AT  14.700 PSIA
                  FLUE GAS                          ENTRAINED FLY ASH
         FLOW     3.3194+05 ACFM           FLOW     2.5200-01 LB/MIN
         SO2      6.0600-02 MOLE %         CAO      11.200     WT %
         CO2      1.1100+01 MOLE %         CASO4     5.700     WT %
         NO       0.0000    MOLE %         MGO       2.500     WT %
         NO2      0.0000    MOLE %         NA2O       .000     WT %
         O2       1.1040+01 MOLE %         NACL      0.000     WT %
         SO3      0.0000    MOLE %         INERT    78.300     WT %
         N2       6.3730+01 MOLE %
         HCL      0.0000    MOLE %
         H2O      1.7100+01 MOLE %

    WTRMKP, EQUIPMENT NUMBER  3 *   70. DEG F
         SO3=     0.0000    MG/L            CA++     7.2100+01 MG/L
         CO3=     7.6300+01 MG/L            MG++     0.0000    MG/L
         SO4=     7.9000+01 MG/L            NA+      0.0000    MG/L
         NO3=     0.0000+00 MG/L            CL-      2.0000+01 MG/L
```

```
22 JUN 76     11:24:59.810                    SINTER PLANT ( 39% RECYCLE ) LIMESTONE SCRUBBING SYSTEM
 STREAM NUMBER                   1             2            3             4            5             6
 TOTAL STREAM
    FLOW RATE (G/SEC)         456.17        .16500+06    1114.6       .16489+06    1675.0       .19458+07
              (G-MOLE/SEC)    4.6014        5780.2       61.857       5779.5       60.278       96469.
    TEMPERATURE (DEG. K)      330.24        330.38       294.27       330.24       330.24       330.24
    PRESSURE  (ATM)           .00000        1.0003       .00000       1.0003       .00000       .00000
    ENTHALPY  (CAL/SEC)     -.13145+07    -.11635+09   -.42306+07   -.11647+09   -.56096+07   -.70715+10
    SOLIDS    (WT %)        100.000       .12004-02    .00000       .36015-03    40.000       11.981
              (MG/CU M)      .00000        12.644       .00000       3.7930       .00000       .00000
 GAS PHASE
    FLOW RATE (G/SEC)        .00000        .16499+06    .00000       .16489+06    .00000       .00000
              (G-MOLE/SEC)   .00000        5780.2       .00000       5779.5       .00000       .00000
              (L/SEC AT T)   .00000        .15666+06    .00000       .15657+06    .00000       .00000
              (L/SEC, STP)   .00000        .12956+06    .00000       .12954+06    .00000       .00000
    ENTHALPY  (CAL/SEC)      .00000       -.11634+09    .00000      -.11647+09    .00000       .00000
    DENSITY   (G/L AT T)     .00000        1.0532       .00000       1.0532       .00000       .00000
    COMP      (G-MOLES/SEC)
      SO2                    .00000        4.0230       .00000       .38621       .00000       .00000
      CO2                    .00000        641.60       .00000       645.23       .00000       .00000
      NO                     .00000        .00000       .00000       .00000       .00000       .00000
      NO2                    .00000        .00000       .00000       .00000       .00000       .00000
      O2                     .00000        635.82       .00000       634.55       .00000       .00000
      SO3                    .00000        .00000       .00000       .00000       .00000       .00000
      N2                     .00000        3510.3       .00000       3510.3       .00000       .00000
      HCL                    .00000        .00000       .00000       .00000       .00000       .00000
      H2O                    .00000        988.41       .00000       989.03       .00000       .00000
      CH4                    .00000        .00000       .00000       .00000       .00000       .00000
    COMP      (MOLE %)
      SO2                    .00000        .69600-01    .00000       .66824-02    .00000       .00000
      CO2                    .00000        11.100       .00000       11.164       .00000       .00000
      NO                     .00000        .00000       .00000       .00000       .00000       .00000
      NO2                    .00000        .00000       .00000       .00000       .00000       .00000
      O2                     .00000        11.000       .00000       10.979       .00000       .00000
      SO3                    .00000        .00000       .00000       .00000       .00000       .00000
      N2                     .00000        60.730       .00000       60.737       .00000       .00000
      HCL                    .00000        .00000       .00000       .00000       .00000       .00000
      H2O                    .00000        17.100       .00000       17.113       .00000       .00000
      CH4                    .00000        .00000       .00000       .00000       .00000       .00000
```

```
22 JUN 76    11:25:00.037                            SINTER PLANT ( 39% RECYCLE ) LIMESTONE SCRUBBING SYSTEM
 STREAM NUMBER                          1           2           3           4           5           6
 SLURRY STREAM
   FLOW RATE  (G/SEC)                .00000      .00000      .00000      .00000      1675.0      .19458+07
              (G-MOLE/SEC)           .00000      .00000      .00000      .00000      68.278      96469.
              (L/SEC AT T)           .00000      .00000      .00000      .00000      1.0183      1734.2
   ENTHALPY   (CAL/SEC)              .00000      .00000      .00000      .00000     -.56096+07  -.70715+10
   BULK DENSITY(G/ML AT T)           .00000      .00000      .00000      .00000      1645.0      1122.0
   KEY COMP RATE(GMOLES/SEC)
     SO2                             .00000      .00000      .00000      .00000      1.0910      380.81
     CO2                             .00000      .00000      .00000      .00000      .79089      280.42
     SO3                             .00000      .00000      .00000      .00000      2.5476      910.97
     V2O5                            .00000      .00000      .00000      .00000      .18004-04   .30669-01
     CAO                             .00000      .00000      .00000      .00000      4.4264      1567.0
     MGO                             .00000      .00000      .00000      .00000      .89429-03   1.5244
     NA2O                            .00000      .00000      .00000      .00000      .94094-03   1.6036
     HCL                             .00000      .00000      .00000      .00000      .25140-02   4.2846
     H2O                             .00000      .00000      .00000      .00000      61.239      96738.
 LIQUID PHASE
   FLOW RATE  (G/SEC)                .00000      .00000      1114.6      .00000      1005.0      .17127+07
              (G-MOLES/SEC)          .00000      .00000      61.857      .00000      55.676      94867.
   ENTHALPY   (CAL/SEC)              .00000      .00000     -.42306+07   .00000     -.37754+07  -.64334+10
   DENSITY    (G/ML AT T)            .00000      .00000      .99826      .00000      .98742      .98759
   KEY COMP RATE(G-MOLES/SEC)
     SO2                             .00000      .00000      .00000      .00000      .43480-03   1.0353
     CO2                             .00000      .00000      .14198-02   .00000      .35443-02   5.8335
     SO3                             .00000      .00000      .92754-03   .00000      .17082-01   31.201
     V2O5                            .00000      .00000      .18006-04   .00000      .18004-04   .30669-01
     CAO                             .00000      .00000      .20006-02   .00000      .17995-01   32.848
     MGO                             .00000      .00000      .00000      .00000      .89429-03   1.5244
     NA2O                            .00000      .00000      .00000      .00000      .94094-03   1.6036
     HCL                             .00000      .00000      .90063-03   .00000      .25140-02   4.2846
     H2O                             .00000      .00000      61.852      .00000      55.633      94789.
   KEY COMP CONC(G-MOLES/KG H2O)
     SO2                             .00000      .00000      .00000      .00000      .43382-03   .60028-03
     CO2                             .00000      .00000      .12742-02   .00000      .35363-02   .34161-02
     SO3                             .00000      .00000      .83239-03   .00000      .17043-01   .18271-01
     V2O5                            .00000      .00000      .16159-04   .00000      .17963-04   .17959-04
     CAO                             .00000      .00000      .18026-02   .00000      .17954-01   .19235-01
     MGO                             .00000      .00000      .00000      .00000      .89228-03   .89269-03
     NA2O                            .00000      .00000      .00000      .00000      .93882-03   .93907-03
     HCL                             .00000      .00000      .80824-03   .00000      .25083-02   .25090-02
   PH                                .00000      .00000      7.1250      .00000      6.3878      6.3397
```

```
                                               SINTER PLANT ( 39% RECYCLE ) LIMESTONE SCRUBBING SYSTEM
22 JUN 76    11:25:00.283         1            2            3            4            5            6
 STREAM NUMBER
 LIQUID PHASE(CONT)
    LIQ. H2O RATE (KG/SEC)     .00000       .00000       1.1143       .00000       1.0023       1707.7
               (LITERS/SEC)    .00000       .00000       1.1163       .00000       1.0150       1729.1
   CONC          (G-MOLES/KG H2O)
     H+                        .00000       .00000       .80772-07    .00000       .48348-06    .54125-06
     OH-                       .00000       .00000       .10804-06    .00000       .24654-06    .22158-06
     HSO3-                     .00000       .00000       .00000       .00000       .16783-03    .24580-03
     SO3=                      .00000       .00000       .00000       .00000       .29356-04    .38968-04
     SO4=                      .00000       .00000       .71125-03    .00000       .10245-01    .10901-01
     HCO3-                     .00000       .00000       .10734-02    .00000       .19976-02    .18489-02
     CO3=                      .00000       .00000       .78269-06    .00000       .64386-06    .54813-06
     NO3-                      .00000       .00000       .32247-04    .00000       .35039-04    .34997-04
     HSO4-                     .00000       .00000       .35673-08    .00000       .54464-06    .63723-06
     H2SO3(L)                  .00000       .00000       .00000       .00000       .80658-08    .13135-07
     H2CO3(L)                  .00000       .00000       .17608-03    .00000       .13000-02    .13379-02
     CA++                      .00000       .00000       .16574-02    .00000       .11033-01    .11644-01
     CAOH+                     .00000       .00000       .30085-08    .00000       .35488-07    .33246-07
     CASO3(L)                  .00000       .00000       .00000       .00000       .23144-03    .31487-03
     CACO3(L)                  .00000       .00000       .10403-05    .00000       .31530-05    .27108-05
     CAHCO3+                   .00000       .00000       .22899-04    .00000       .22576-03    .21776-03
     CASO4(L)                  .00000       .00000       .12114-03    .00000       .64898-02    .70547-02
     CANO3+                    .00000       .00000       .79453-07    .00000       .87159-06    .90536-06
     MG++                      .00000       .00000       .00000       .00000       .62131-03    .61581-03
     MGOH+                     .00000       .00000       .00000       .00000       .40824-07    .35879-07
     MGSO3(L)                  .00000       .00000       .00000       .00000       .51892-05    .66232-05
     MGHCO3+                   .00000       .00000       .00000       .00000       .63275-05    .57258-05
     MGSO4(L)                  .00000       .00000       .00000       .00000       .25912-03    .26425-03
     MGCO3(L)                  .00000       .00000       .00000       .00000       .29918-06    .24131-06
     NA+                       .00000       .00000       .00000       .00000       .18265-02    .18251-02
     NAOH(L)                   .00000       .00000       .00000       .00000       .80559-10    .71775-10
     NACO3-                    .00000       .00000       .00000       .00000       .52560-07    .43355-07
     NAHCO3(L)                 .00000       .00000       .00000       .00000       .23909-05    .21933-05
     NASO4-                    .00000       .00000       .00000       .00000       .48706-04    .50796-04
     NANO3(L)                  .00000       .00000       .00000       .00000       .16207-07    .16011-07
     CL-                       .00000       .00000       .80824-03    .00000       .25083-02    .25090-02
   IONIC STRENGTH              .00000       .00000       .57974-02    .00000       .47208-01    .49771-01
   DISSOLVED SOLIDS (WPPM)     .00000       .00000       264.76       .00000       2757.3       2932.2
```

```
22 JUN 76    11:25:00.542                      SINTER PLANT ( 39% RECYCLE ) LIMESTONE SCRUBBING SYSTEM
 STREAM NUMBER                 1           2           3           4           5           6
 SOLID PHASE
   FLOW RATE  (G/SEC)        456.17      1.9807      .00000      .59380      670.02      .23312+06
              (G-MOLE/SEC)   4.6014      .27294-01   .00000      .81833-02   4.6019      1601.5
   ENTHALPY   (CAL/SEC)     -.13145+07  -5893.4      .00000     -1767.0     -.18342+07  -.63816+09
   SPECIFIC GRAVITY          2.7271      3.1515      .00000      3.1515      2.4226      2.4228
   KEY COMP RATE(GMOLES/SEC)
     SO2                     .00000      .00000      .00000      .00000      1.0906      379.78
     CO2                     4.4208      .00000      .00000      .00000      .78734      274.59
     SO3                     .00000      .12657-02   .00000      .37949-03   2.5305      879.77
     N2O5                    .00000      .00000      .00000      .00000      .00000      .00000
     CAO                     4.4208      .51508-02   .00000      .15443-02   4.4084      1534.1
     MGO                     .00000      .12772-02   .00000      .38294-03   .00000      .00000
     NA2O                    .00000      .13439-02   .00000      .40294-03   .00000      .00000
     HCL                     .00000      .23044-02   .00000      .69091-03   .00000      .00000
     H2O                     .00000      .00000      .00000      .00000      5.6063      1949.4
   COMP RATE     (G MOLES/SEC)
     CAO                     .00000      .38851-02   .00000      .11648-02   .00000      .00000
     CA(OH)2                 .00000      .00000      .00000      .00000      .00000      .00000
     CACO3                   4.4208      .00000      .00000      .00000      .78734      274.59
     CASO3                   .00000      .00000      .00000      .00000      .00000      .00000
     CASO3*1/2H2O            .00000      .00000      .00000      .00000      1.0906      379.78
     CASO4                   .00000      .12657-02   .00000      .37949-03   .00000      .10891-02
     CASO4*2H2O              .00000      .00000      .00000      .00000      2.5305      879.77
     MGO                     .00000      .12772-02   .00000      .38294-03   .00000      .00000
     MG(OH)2                 .00000      .00000      .00000      .00000      .00000      .00000
     MGCO3                   .00000      .00000      .00000      .00000      .00000      .00000
     MGCO3*3H2O              .00000      .00000      .00000      .00000      .00000      .00000
     MGCO3*5H2O              .00000      .00000      .00000      .00000      .00000      .00000
     MGSO3                   .00000      .00000      .00000      .00000      .00000      .00000
     MGSO3*3H2O              .00000      .00000      .00000      .00000      .00000      .00000
     MGSO3*6H2O              .00000      .00000      .00000      .00000      .00000      .00000
     MGSO4                   .00000      .00000      .00000      .00000      .00000      .00000
     MGCL2                   .00000      .00000      .00000      .00000      .00000      .00000
     NA2O                    .00000      .19174-03   .00000      .57487-04   .00000      .00000
     NACL                    .00000      .23044-02   .00000      .69091-03   .00000      .00000
     INERTS                  .18054      .18370-01   .00000      .55077-02   .19340      67.390
     LIMESTONE               .00000      .00000      .00000      .00000      .00000      .00000
   XLS                       .00000      .00000      .00000      .00000      .00000      .00000
```

```
22 JUN 78    11:25:00.775              SINTER PLANT ( 39% RECYCLE ) LIMESTONE SCRUBBING SYSTEM
 STREAM NUMBER               1            2            3            4            5            6
 ELEMENTS(G-ATOMS/SEC
 TOTAL STREAM
       H                  .00000       1976.8       123.70       1978.1       122.48       .19348+06
       C                  4.4208       641.60       .14198-02    645.23       .79089       280.42
       N                  .00000       7020.7       .36012-04    7020.7       .36008-04    .61337-01
       O                  13.262       3551.3       61.859       3549.4       77.074       .10236+06
       NA                 .00000       .26879-02    .00000       .80588-03    .18819-02    3.2073
       MG                 .00000       .12772-02    .00000       .38294-03    .89429-03    1.5244
       S                  .00000       4.0243       .92754-03    .38659       3.6386       1291.8
       CL                 .00000       .23044-02    .90063-03    .69091-03    .25140-02    4.2846
       CA                 4.4208       .51508-02    .20086-02    .15443-02    4.4264       1567.0
 GAS PHASE
       H                  .00000       1976.8       .00000       1978.1       .00000       .00000
       C                  .00000       641.60       .00000       645.23       .00000       .00000
       N                  .00000       7020.7       .00000       7020.7       .00000       .00000
       O                  .00000       3551.3       .00000       3549.4       .00000       .00000
       S                  .00000       4.0230       .00000       .38521       .00000       .00000
       CL                 .00000       .00000       .00000       .00000       .00000       .00000
 LIQUID PHASE
       H                  .00000       .00000       123.70       .00000       111.27       .18958+06
       C                  .00000       .00000       .14198-02    .00000       .35443-02    5.8335
       N                  .00000       .00000       .36012-04    .00000       .36008-04    .61337-01
       O                  .00000       .00000       61.859       .00000       55.712       94932.
       NA                 .00000       .00000       .00000       .00000       .18819-02    3.2073
       MG                 .00000       .00000       .00000       .00000       .89429-03    1.5244
       S                  .00000       .00000       .92754-03    .00000       .17517-01    32.236
       CL                 .00000       .00000       .90063-03    .00000       .25140-02    4.2846
       CA                 .00000       .00000       .20086-02    .00000       .17995-01    32.848
 SOLID PHASE
       H                  .00000       .23044-02    .00000       .69091-03    11.213       3898.9
       C                  4.4208       .00000       .00000       .00000       .78734       274.59
       N                  .00000       .00000       .00000       .00000       .00000       .00000
       O                  13.262       .11569-01    .00000       .34687-02    21.362       7431.6
       NA                 .00000       .26879-02    .00000       .80588-03    .00000       .00000
       MG                 .00000       .12772-02    .00000       .38294-03    .00000       .00000
       S                  .00000       .12657-02    .00000       .37949-03    3.6211       1259.5
       CL                 .00000       .23044-02    .00000       .69091-03    .00000       .00000
       CA                 4.4208       .51508-02    .00000       .15443-02    4.4064       1534.1
```

```
22 JUN 76    11:25:01.855                              SINTER PLANT ( 39% RECYCLE ) LIMESTONE SCRUBBING SYSTEM
 STREAM NUMBER                       7            8            9            10
 TOTAL STREAM
   FLOW RATE (G/SEC)              5614.1      .19460+07    .19514+07    3908.4
             (G-MOLE/SEC)         278.33      96469.       96747.       216.52
   TEMPERATURE (DEG. K)           330.24      330.24       330.24       330.24
   PRESSURE  (ATM)                .00000      .00000       .00000       .00000
   ENTHALPY  (CAL/SEC)         -.20403+08   -.70717+10   -.70919+10   -.14682+08
   SOLIDS    (WT %)               11.981      11.976       11.981       .00000
             (MG/CU M)            .00000      .00000       .00000       .00000
 GAS PHASE
   FLOW RATE (G/SEC)              .00000      .00000       .00000       .00000
             (G-MOLE/SEC)         .00000      .00000       .00000       .00000
             (L/SEC AT T)         .00000      .00000       .00000       .00000
             (L/SEC, STP)         .00000      .00000       .00000       .00000
   ENTHALPY  (CAL/SEC)            .00000      .00000       .00000       .00000
   DENSITY   (G/L AT T)           .00000      .00000       .00000       .00000
   COMP      (G-MOLES/SEC)
     SO2                          .00000      .00000       .00000       .00000
     CO2                          .00000      .00000       .00000       .00000
     NO                           .00000      .00000       .00000       .00000
     NO2                          .00000      .00000       .00000       .00000
     O2                           .00000      .00000       .00000       .00000
     SO3                          .00000      .00000       .00000       .00000
     N2                           .00000      .00000       .00000       .00000
     HCL                          .00000      .00000       .00000       .00000
     H2O                          .00000      .00000       .00000       .00000
     CH4                          .00000      .00000       .00000       .00000
   COMP      (MOLE %)
     SO2                          .00000      .00000       .00000       .00000
     CO2                          .00000      .00000       .00000       .00000
     NO                           .00000      .00000       .00000       .00000
     NO2                          .00000      .00000       .00000       .00000
     O2                           .00000      .00000       .00000       .00000
     SO3                          .00000      .00000       .00000       .00000
     N2                           .00000      .00000       .00000       .00000
     HCL                          .00000      .00000       .00000       .00000
     H2O                          .00000      .00000       .00000       .00000
     CH4                          .00000      .00000       .00000       .00000
```

```
22 JUN 76     11:25:02.036
 STREAM NUMBER                    7                  SINTER PLANT ( 39% RECYCLE ) LIMESTONE SCRUBBING SYSTEM
 SLURRY STREAM                                 8            9            10
    FLOW RATE (G/SEC)          5614.1       .19460+07    .19514+07     .00000
              (G-MOLE/SEC)     278.33       96469.       96747.        .00000
              (L/SEC AT T)     5.0040       1734.2       1739.2        .00000
   ENTHALPY   (CAL/SEC)      -.20403+08   -.70717+10   -.70919+10     .00000
   BULK DENSITY(G/ML AT T)     1121.9       1122.1       1122.0        .00000
   KEY COMP RATE(GMOLES/SEC)
     SO2                       1.0987       381.91       381.91        .00000
     CO2                       .80908       276.80       281.23        .00000
     SO3                       2.6283       913.53       913.60        .00000
     N2O5                      .88485-04    .30670-01    .30757-01     .00000
     CAO                       4.5211       1567.0       1571.5        .00000
     MGO                       .43983-02    1.5253       1.5288        .00000
     NA2O                      .46268-02    1.6047       1.6083        .00000
     HCL                       .12362-01    4.2863       4.2970        .00000
     H2O                       279.11       96739.       97017.        .00000
 LIQUID PHASE
    FLOW RATE (G/SEC)          4941.5       .17129+07    .17176+07     3908.4
              (G-MOLES/SEC)    273.71       94869.       95141.        216.52
   ENTHALPY   (CAL/SEC)      -.18562+08   -.64338+10   -.64519+10   -.14682+08
   DENSITY    (G/ML AT T)      .98759       .98772       .98759        .98742
   KEY COMP RATE(G-MOLES/SEC)
     SO2                       .29871-02    2.0751       1.0383        .16909-02
     CO2                       .16831-01    4.2923       5.8504        .13783-01
     SO3                       .90021-01    33.013       31.291        .66430-01
     N2O5                      .88485-04    .30670-01    .30757-01     .70015-04
     CAO                       .94773-01    34.156       32.943        .69980-01
     MGO                       .43983-02    1.5253       1.5208        .34778-02
     NA2O                      .46268-02    1.6047       1.6083        .36592-02
     HCL                       .12362-01    4.2863       4.2970        .97767-02
     H2O                       273.49       94788.       95462.        216.35
   KEY COMP CONC(G-MOLES/KG H2O)
     SO2                       .60628-03    .12152-02    .60628-03     .43382-03
     CO2                       .34161-02    .25136-02    .34161-02     .35363-02
     SO3                       .18271-01    .19332-01    .18271-01     .17043-01
     N2O5                      .17959-04    .17960-04    .17959-04     .17963-04
     CAO                       .19235-01    .20031-01    .19235-01     .17954-01
     MGO                       .89269-03    .89322-03    .89269-03     .89228-03
     NA2O                      .93907-03    .93969-03    .93907-03     .93882-03
     HCL                       .25090-02    .25100-02    .25090-02     .25083-02
   PH                          6.3397       5.8535       6.3397        6.3878
```

```
22 JUN 76    11:25:02,250                SINTER PLANT ( 39% RECYCLE ) LIMESTONE SCRUBBING SYSTEM
 STREAM NUMBER                   7            8            9            10
 LIQUID PHASE(CONT)
   LIQ, H2O RATE (KG/SEC)     4,9270       1707,7       1712,6       3,8977
              (LITERS/SEC)    4,9889       1728,9       1734,1       3,9473
   CONC        (G-MOLES/KG H2O)
     H+                       ,54125-06    ,16604-05    ,54125-06    ,48348-06
     OH-                      ,22158-06    ,72498-07    ,22158-06    ,24654-06
     HSO3-                    ,24580-03    ,81661-03    ,24580-03    ,16783-03
     SO3=                     ,38968-04    ,42577-04    ,38968-04    ,29356-04
     SO4=                     ,10901-01    ,11485-01    ,10901-01    ,10245-01
     HCO3-                    ,18489-02    ,75341-03    ,18489-02    ,19976-02
     CO3=                     ,54013-06    ,72386-07    ,54013-06    ,64386-06
     NO3-                     ,34997-04    ,34979-04    ,34997-04    ,35039-04
     HSO4-                    ,63723-06    ,20372-05    ,63723-06    ,54464-06
     H2SO3(L)                 ,13135-07    ,13332-06    ,13135-07    ,80658-08
     H2CO3(L)                 ,13379-02    ,16656-02    ,13379-02    ,13000-02
     CA++                     ,11644-01    ,12039-01    ,11644-01    ,11003-01
     CAOH+                    ,33246-07    ,11143-07    ,33246-07    ,35488-07
     CASO3(L)                 ,31487-03    ,34879-03    ,31487-03    ,23144-03
     CACO3(L)                 ,27108-05    ,36832-06    ,27108-05    ,31530-05
     CAHCO3+                  ,21776-03    ,90888-04    ,21776-03    ,22576-03
     CASO4(L)                 ,70547-02    ,75211-02    ,70547-02    ,64898-02
     CANO3+                   ,90536-06    ,92566-06    ,90536-06    ,87159-06
     MG++                     ,61581-03    ,61251-03    ,61581-03    ,62131-03
     MGOH+                    ,35879-07    ,11581-07    ,35879-07    ,40824-07
     MGSO3(L)                 ,66232-05    ,70655-05    ,66232-05    ,51892-05
     MGHCO3+                  ,57258-05    ,23014-05    ,57258-05    ,63275-05
     MGSO4(L)                 ,26425-03    ,27130-03    ,26425-03    ,25912-03
     MGCO3(L)                 ,24131-06    ,31575-07    ,24131-06    ,29918-06
     NA+                      ,18251-02    ,18256-02    ,18251-02    ,18265-02
     NAOH(L)                  ,71775-10    ,23376-10    ,71775-10    ,80559-10
     NACO3-                   ,43355-07    ,57547-08    ,43355-07    ,52560-07
     NAHCO3(L)                ,21933-05    ,88954-06    ,21933-05    ,23909-05
     NASO4-                   ,50796-04    ,52905-04    ,50796-04    ,48706-04
     NANO3(L)                 ,16011-07    ,15907-07    ,16011-07    ,16207-07
     CL-                      ,25090-02    ,25100-02    ,25090-02    ,25083-02
   IONIC STRENGTH             ,49771-01    ,51404-01    ,49771-01    ,47208-01
   DISSOLVED SOLIDS (WPPM)    2932,2       3058,8       2932,2       2757,3
```

```
22 JUN 76    11:25:02,475                  SINTER PLANT ( 39% RECYCLE ) LIMESTONE SCRUBBING SYSTEM
 STREAM NUMBER                   7             8              9              10
 SOLID PHASE
   FLOW RATE (G/SEC)           672,60      ,23305+06      ,23379+06      ,00000
             (G-MOLE/SEC)      4,6207      1600,3         1606,1         ,00000
   ENTHALPY  (CAL/SEC)       -,18412+07  -,63794+09     -,64000+09      ,00000
   SPECIFIC GRAVITY            2,4228      2,4225         2,4228         ,00000
   KEY COMP RATE(GMOLES/SEC)
     SO2                       1,0957      379,83         380,87         ,00000
     CO2                       ,79225      272,50         275,38         ,00000
     SO3                       2,5383      880,52         882,31         ,00000
     N2O5                      ,00000      ,00000         ,00000         ,00000
     CAO                       4,4263      1532,9         1538,6         ,00000
     MGO                       ,00000      ,00000         ,00000         ,00000
     NA2O                      ,00000      ,00000         ,00000         ,00000
     HCL                       ,00000      ,00000         ,00000         ,00000
     H2O                       5,6245      1951,0         1955,1         ,00000
   COMP RATE    (G MOLES/SEC)
     CAO                       ,00000      ,00000         ,00000         ,00000
     CA(OH)2                   ,00000      ,00000         ,00000         ,00000
     CACO3                     ,79225      272,50         275,38         ,00000
     CASO3                     ,00000      ,00000         ,00000         ,00000
     CASO3*1/2H2O              1,0957      379,83         380,87         ,00000
     CASO4                     ,29114-05   ,10120-02      ,10120-02      ,00000
     CASO4*2H2O                2,5383      880,52         882,31         ,00000
     MGO                       ,00000      ,00000         ,00000         ,00000
     MG(OH)2                   ,00000      ,00000         ,00000         ,00000
     MGCO3                     ,00000      ,00000         ,00000         ,00000
     MGCO3*3H2O                ,00000      ,00000         ,00000         ,00000
     MGCO3*5H2O                ,00000      ,00000         ,00000         ,00000
     MGSO3                     ,00000      ,00000         ,00000         ,00000
     MGSO3*3H2O                ,00000      ,00000         ,00000         ,00000
     MGSO3*6H2O                ,00000      ,00000         ,00000         ,00000
     MGSO4                     ,00000      ,00000         ,00000         ,00000
     MGCL2                     ,00000      ,00000         ,00000         ,00000
     NA2O                      ,00000      ,00000         ,00000         ,00000
     NACL                      ,00000      ,00000         ,00000         ,00000
     INERTS                    ,19443      67,403         67,584         ,00000
     LIMESTONE                 ,00000      ,00000         ,00000         ,00000
   XLS                         ,00000      ,00000         ,00000         ,00000
```

```
22 JUN 76    11:25:02.065                     SINTER PLANT ( 39% RECYCLE ) LIMESTONE SCRUBBING SYSTEM
 STREAM NUMBER                       7           8           9          10
 ELEMENTS(G-ATOMS/SEC
 TOTAL STREAM
      H                        558.23      .19348+06   .19404+06   432.71
      C                        .80908      276.80      281.23      .13783-01
      N                        .17697-03   .61341-01   .61514-01   .14003-03
      O                        295.34      .10237+06   .10266+06   216.66
      NA                       .92536-02   3.2093      3.2165      .73184-02
      MG                       .43983-02   1.5253      1.5288      .34778-02
      S                        3.7271      1295.4      1295.5      .68120-01
      CL                       .12362-01   4.2863      4.2970      .97767-02
      CA                       4.5211      1567.0      1571.5      .69980-01
 GAS PHASE
      H                        .00000      .00000      .00000      .00000
      C                        .00000      .00000      .00000      .00000
      N                        .00000      .00000      .00000      .00000
      O                        .00000      .00000      .00000      .00000
      S                        .00000      .00000      .00000      .00000
      CL                       .00000      .00000      .00000      .00000
 LIQUID PHASE
      H                        546.98      .18958+06   .19013+06   432.71
      C                        .16831-01   4.2923      5.8504      .13783-01
      N                        .17697-03   .61341-01   .61514-01   .14003-03
      O                        273.90      94937.      95206.      216.66
      NA                       .92536-02   3.2093      3.2165      .73184-02
      MG                       .43983-02   1.5253      1.5288      .34778-02
      S                        .93008-01   35.088      32.329      .68120-01
      CL                       .12362-01   4.2863      4.2970      .97767-02
      CA                       .94773-01   34.156      32.943      .69980-01
 SOLID PHASE
      H                        11.249      3901.9      3910.1      .00000
      C                        .79225      272.50      275.38      .00000
      N                        .00000      .00000      .00000      .00000
      O                        21.442      7430.0      7453.0      .00000
      NA                       .00000      .00000      .00000      .00000
      MG                       .00000      .00000      .00000      .00000
      S                        3.6341      1260.4      1263.2      .00000
      CL                       .00000      .00000      .00000      .00000
      CA                       4.4263      1532.9      1538.6      .00000
```

```
22 JUN 76    11:24:59.710                    SINTER PLANT ( 39% RECYCLE ) LIMESTONE SCRUBBING SYSTEM

  SCRUB5, EQUIPMENT NUMBER  7
       L/G = 8.15413+01 GAL/1000 ACF
       PRESSURE DROP =    .00 PSI
       PARTICULATE REMOVAL =  70.02 %
       KGAV(SO2) =  1.07662+05 LBMOLE/HR ATM
       NTU(SO2) =  2.347

  STOICHIOMETRY = 109.9 MOLE % LIMESTONE PER SO2
```

APPENDIX D

COST DATA

1.0 EQUIPMENT LIST

The items listed in Tables D-1 and D-2 are the equipment required to process the flue gas from the two sinter plant cases considered in this study. The equipment arrangement was shown in Figure 5-1. The limestone process was divided into ten areas to allow comparisons of the two cases to be made easily.

The size cost scale factors listed in Tables D-1 and D-2 were used to obtain order-of-magnitude cost estimates for the process equipment. Capital investment costs for complete plants can be correlated to within ± 30% with these factors using some plant parameter as a basis for equipment sizing. The correlation used in this study was of the form:

$$\text{Capital Cost} = \left[\begin{array}{c}\text{Capital cost}\\\text{for reference}\\\text{size}\end{array}\right] \times \left[\frac{\text{size parameter}}{\text{reference size parameter}}\right]^{n} \times I$$

where: I = inflation index factor
n = size-cost scale factor

This technique allows a reasonably accurate estimate of plant cost to be made from data on different size plants without obtaining price quotations from equipment vendors.

The inflation indices used for cost estimates of this sort introduce some error into the estimate, but this is unavoidable. Several publications, most notably Marshall and Stevens, Chemical Engineering Magazine, and Nelson Refinery Construction and Equipment Inflation Indices, list inflation indices based on a year in the late 1950's (e.g., 1957 = 100.). Using these indices, costs of equipment can be scaled up to the present with reasonable accuracy. However, projecting costs into the future (1977 in this case) introduces a possible error. Normally, some inflation rate in the 7-9 percent per year range is used to estimate future equipment prices. The inflation rate assumed can be based on past performance as indicated by the published inflation indices but projecting past inflation rates into the future can introduce error.

The Chemical Engineering plant cost index was used for this study. The 1974 costs were scaled up to 1975 costs using indices from this index. To project costs from 1975 to 1977 a yearly inflation rate of 7 percent was used based upon inflation rates of previous years. The high inflation rates experienced since 1973 were considered abnormally high so they were not used.

TABLE D-1

CASE 1: STANDARD OPERATION

WORK SHEET FOR PROCESS EQUIPMENT COSTS

AREA 1 - MATERIALS HANDLING

Item	No.	Description	Size-Cost Scale Factor	Factor Source	Base Cost Each (1977)	Total Mid-1977 Cost
1. Unloading hopper No. 1	1	Capacity .31m^3, carbon steel	0.68	Chem. Engr. 3-24-69 Guthrie	560	560
2. Limestone feeder No. 1 (vibrating)	1	5.8 kg/s	0.58	Chem. Engr. 3-24-69 Guthrie	1,100	1,100
3. Conveyor (belt) No. 1	1	5.8 kg/s	0.81 0.65	Fund. of Cost Engr. 1964 Chem. Engr. 3-24-69 Guthrie	580	580
4. Conveyor (belt) No. 2	1	5.8 kg/s	0.81 0.65	Fund. of Cost Engr. 1964 Chem. Engr. 3-24-69 Guthrie	2,750	2,750
5. Hoppers under pile	3	Capacity 0.21m^3, carbon steel	0.68	Chem. Engr. 3-24-69 Guthrie	470	1,410
6. Limestone feeder No. 2 (vibrating)	3	2.8 kg/s	0.58	Chem. Engr. 3-24-69 Guthrie	570	1,710
7. Conveyor (belt) No. 3	1	2.8 kg/s	0.65 0.81	Chem. Engr. 3-24-69 Guthrie Fund. of Cost Engr. 1964	4,240	4,240
8. Tunnel sump pump	2	$3.2 \times 10^{-4} m^3/s$, carbon steel, neoprene lining, 186.5 watt motor	Depends on gpm and head requirements resulting in changes of motor and impeller size		720	1,440
9. Elevator No. 1	1	2.8 kg/s	0.83	Chem. Engr. 3-24-69	2,560	2,560
10. Bin	1	Capacity 15.6m^3, carbon steel	0.68	Chem. Engr. 3-24-69	4,280	4,280
11. Car shaker	1	Railroad trackside vibrator	----	----	7,950	7,950
12. Dust collecting system No. 1	1	0.12 m^3/s inertial separator, cyclone, hoppers, fan, and drive	0.80	Chem. Engr. 3-24-69 Guthrie	590	590
13. Dust collecting system No. 2	1	0.33 m^3/s inertial separator, cyclone, hoppers, fan, and drive	0.80	Chem. Engr. 3-24-69	1,100	1,100
14. Bag filter system	1	0.87 m^3/s, automatic fabric dust collectors, bag support, shaker system, isolation damper, motor, drive, dust hopper, fan and motor	0.68	Chem. Engr. 3-24-69	2,580	2,580
SUBTOTAL						32,850

TABLE D-1 (Continued)

AREA 2 - FEED PREPARATION

Item	No.	Description	Size-Cost Scale Factor	Factor Source	Base Cost Each (1977)	Total Mid-1977 Cost
1. Bin discharge feeder	1	0.8 kg/s, carbon steel	0.58	Chem. Engr. 3-24-69 Guthrie	320	320
2. Weigh feeder	1	0.8 kg/s, carbon steel	0.65	Chem. Engr. 3-24-69 Guthrie	3,900	3,900
3. Gyratory crusher	1	0.8 kg/s	1.20	Chem. Engr. 3-24-69 Guthrie	2,450	2,450
4. Elevator No. 2	1	0.8 kg/s	0.65	Chem. Engr. 3-24-69 Guthrie	1,140	1,140
5. Wet ball mill	1	7.4 kg/s	0.65	Chem. Engr. 3-24-69 Guthrie	50,550	50,550
	1	78330 W motor	1.07	Fund. of Cost Engr. 1964	3,600	3,600
6. Slurry feed tank	1	Capacity 20.8m^3, carbon steel	0.68	Chem. Engr. 3-24-69 Guthrie	5,450	5,450
Lining	1	6.35×10^{-3}m neoprene	----	----	4,820	4,820
7. Agitator, slurry feed tank	1	1492 W, neoprene coated	0.50	Chem. Engr. 3-24-69 Guthrie	3,020	3,020
			0.46	Fund. of Cost Engr. 1964		
8. Pumps, slurry feed tank	2	6.9×10^{-4}m^3/s, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		1,990	3,980
9. Dust collecting system	1	0.42 m^3/s, inertial separator, cyclone, hoppers, fan and drive	0.80	Chem. Engr. 3-24-69 Guthrie	1,340	1,340
10. Hoist	1	1800 kg electric	0.81	Popper, H.	10,890	10,890
11. Bag filter system	1	0.87 m^3/s, automatic fabric dust collectors, bag support, shaker system, isolation damper, motor, drive, dust hopper, fan and motor	0.68	Chem. Engr. 3-24-69	2,580	2,580
SUBTOTAL						94,040

TABLE D-1 (Continued)
AREA 3 - PARTICULATE SCRUBBING

Item	No.	Description	Size-Cost Scale Factor	Factor Source	Base Cost Each (1977)	Total Mid-1977 Cost
1. Tank, particulate scrubber, effluent hold	2	Capacity 174.1 m^3, carbon steel	0.68	Chem. Engr. 3-24-69 Guthrie	29,150	58,300
Lining	2	6.35×10^{-3} m neoprene	----	----	21,460	42,920
2. Agitator, effluent hold tank	2	7460 W, neoprene coated	0.26 0.50	Fund. of Cost Engr. 1964 Chem. Engr. 3-24-69 Guthrie	5,770	11,450
3. Pumps, recycle slurry	3	.4 m^3/s, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		20,950	62,850
4. Venturi scrubber	2	93.7 m^3/s, carbon steel, neoprene lined	0.60	Universal Oil Products	160,250	320,500
5. Venturi sump	2	Carbon steel, neoprene lining	0.68	Chem. Engr. 3-24-69 Guthrie	61,430	122,860
6. Soot blowers	10	----	1.00	TVA	4,820	48,200
7. Bleed pump	3	1.9×10^{-3} m^3/s, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		2,000	6,000
SUBTOTAL						673,170

TABLE D-1 (Continued)
AREA 4 - SO_2 SCRUBBING

Item	No.	Description	Size-Cost Scale Factor	Factor Source	Base Cost Each (1977)	Total Mid-1977 Cost
1. Spray tower scrubber	2	Gas flow 93.7 m^3/s, carbon steel, neoprene	---	Western Precipitation Div. Joy Mfr. Co.[a]	232,000	464,000
2. Spray tower sump	2	Carbon steel, neoprene lined	0.68	Chem. Engr. 3-24-69 Guthrie	60,460	120,920
3. Tank absorber effluent hold	2	Capacity 530.9 m^3, carbon steel, field erected	0.68	Chem. Engr. 3-24-69 Guthrie	37,680	75,360
Lining	2	6.35×10^{-3} m neoprene	---	---	32,250	64,500
4. Agitator, SO_2 absorber hold tank	2	22380 W, neoprene coated	0.50	Chem. Engr. 3-24-69 Guthrie	12,360	24,720
5. Pumps, SO_2 absorber recycle slurry	5	.62 m^3/s, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		32,620	163,100
6. Pumps, makeup water	2	1.1×10^{-3} m^3/s, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		1,500	3,000
7. Soot blowers	10		1.00	TVA	4,820	48,200
8. Demister	2	Carbon steel, neoprene lined	----	-----	23,200	46,200
9 Pump, bleed	4	6.7×10^{-4} m^3/s, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		2,000	8,000
10. Tank Demister Wash	2	Capacity 1.89m^3, carbon steel, neoprene lined	0.68	Chem. Engr. 3-24-69 Guthrie	1,400	2,800
11. Pump, Demister Wash	4	1.3×10^{-3} m^3/s, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		1,500	6,000
SUBTOTAL						1,026,800

---

[a] Indicateds source of spray tower cost

TABLE D-1 (Continued)

AREA 5 - REHEAT

Item	No.	Description	Size-Cost Scale Factor	Factor Source	Base Cost Each (1977)	Total Mid-1977 Cost
1. Steam reheater	2	4.0×10^6 W rating $146.3\ m^2$ surface area	0.80	Chem. Engr. 3-24-69 Guthrie	81,130	162,270
2. Soot blowers	10	---	1.00	TVA	4,820	48,200
SUBTOTAL						210,470

AREA 6 - GAS HANDLING

1. Fan	2	1.53×10^5 W motor drive	0.68	Chem. Engr. 3-24-69 Guthrie	71,650	143,300

AREA 7 - SOLIDS DISPOSAL

Item	No.	Description	Size-Cost Scale Factor	Factor Source	Base Cost Each (1977)	Total Mid-1977 Cost
1. Clarifer	1	$5.5 \times 10^{-3}\ m^3/s$	---	PEDCO (PE-146)	161,000	161,000
2. Pumps, pond feed	2	$1.3 \times 10^{-3}\ m^3/s$, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		1,500	3,000
3. Pump, clarifer water recycle	2	$4.1 \times 10^{-3}\ m^3/s$, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		3,500	7,000
4. Pumps, particulate pond water recycle	2	$3.2 \times 10^{-3}\ m^3/s$, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		2,500	5,000
5. Pumps, SO_2 pond water recycle	2	$5.8 \times 10^{-4}\ m^3/s$, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		1,100	2,200
SUBTOTAL						178,200

TABLE D-1 (Continued)

AREA 8 - UTILITIES

Note: There is no process equipment in this area.

AREA 9 - SERVICES

Item	No.	Description	Size-Cost Scale Factor	Factor Source	Base Cost Each (1977)	Total Mid-1977 Cost
1. Payloader			---	---	29,850	29,850
2. Plant vehicles			---	---	---	12,050
3. Maint. & instrument shop-equipment			---	---	31,780	31,780
4. Service building-equipment			---	---	42,130	42,130
5. Stores-equipment			---	---	12,760	12,760
SUBTOTAL						128,570

AREA 10 - PARTICLE RECIRCULATION

Item	No.	Description	Size-Cost Scale Factor	Factor Source	Base Cost Each (1977)	Total Mid-1977 Cost
1. Wet ball mill	1	3.2×10^{-4} m^3/s	.65	McGlamery	29,950	29,950
2. Pump, particle recirculation	2	3.2×10^{-4} m^3/s, molded polypropylane	Depends on gpm and head requirements resulting in changes of motor and impeller sizes		500	1,000
3. Tank, particle recirculation surge	1	Capacity 1.1m^3, carbon steel, neoprene lined	.68	McGlamery	1,000	1,000
SUBTOTAL						31,950

TABLE D-2

CASE 2: 39% SINTER GAS RECYCLE

WORK SHEET FOR PROCESS EQUIPMENT COSTS

AREA 1 - MATERIALS HANDLING

Item	No.	Description	Size-Cost Scale Factor	Factor Source	Base Cost Each (1977)	Total Mid-1977 Cost
1. Unloading hopper No. 1	1	Capacity .34 m^3, carbon steel	0.68	Chem. Engr. 3-24-69 Guthrie	600	600
2. Limestone feeder No. 1 (vibrating)	1	6.3 kg/s	0.58	Chem. Engr. 3-24-69 Guthrie	1,160	1,160
3. Conveyor (belt) No. 1	1	6.3 kg/s	0.81 0.65	Fund. of Cost Engr. 1964 Chem. Engr. 3-24-69 Guthrie	620	620
4. Conveyor (belt) No. 2	1	6.3 kg/s	0.81 0.65	Fund. of Cost Engr. 1964 Chem. Engr. 3-24-69 Guthrie	2,940	2,940
5. Hoppers under pile	3	Capacity 0.23 m^3, carbon steel	0.68	Chem. Engr. 3-24-69 Guthrie	490	1,470
6. Limestone feeder No. 2 (vibrating)	3	3.0 kg/s	0.58	Chem. Engr. 3-24-69	600	1,800
7. Conveyor (belt) No. 3	1	3.0 kg/s	0.65 0.81	Chem. Engr. 3-24-69 Guthrie Fund. of Cost Engr. 1964	4,490	13,470
8. Tunnel sump pump	2	3.2×10^{-4} m^3/s, carbon steel, neoprene lining, 186.5 watt motor	Depends on gpm and head requirements resulting in changes of motor and impeller sizes		720	1,440
9. Elevator No. 1	1	3.0 kg/s	0.83	Chem. Engr. 3-24-69	2,760	2,760
10. Bin	1	Capacity 17 m^2, carbon steel	0.68	Chem. Engr. 3-24-69	5,470	5,470
11. Car shaker	1	Railroad trackside vibrator	----	----	6,600	6,600
12. Dust collecting system No. 1	1	0.12 m^3/s, inertial separators, cyclone, hoppers, fan, and drive	0.80	Chem. Engr. 3-24-69 Guthrie	590	590
13. Dust collecting system No. 2	1	0.35 m^3/s, inertial separators, cyclone, hoppers, fan, and drive	0.80	Chem. Engr. 3-24-69	1,160	1,160
14. Bag filter system	1	0.94 m^3/s, automatic fabric dust collectors, bag support, shaker system, isolation damper, motor, drive, dust hopper, fan and motor	0.68	Chem. Engr. 3-24-69	2,730	2,730
SUBTOTAL						36,210

TABLE D-2 (Continued)

AREA 2 - FEED PREPARATION

Item	NO.	Description	Size-Cost Scale Factor	Factor Source	Base Cost Each (1977)	Total Mid-1977 Cost
1. Bin discharge feeder	1	0.8 kg/s, carbon steel	0.58	Chem. Engr. 3-24-69 Guthrie	320	320
2. Weigh feeder	1	0.8 kg/s, carbon steel	0.65	Chem. Engr. 3-24-69 Guthrie	3,900	3,900
3. Gyratory crusher	1	0.8 kg/s	1.20	Chem. Engr. 3-24-69 Guthrie	2,450	2,450
4. Elevator No. 2	1	0.8 kg/s	0.65	Chem. Engr. 3-24-69 Guthrie	1,140	1,140
5. Wet ball mill	1	7.9 kg/s	0.65	Chem. Engr. 3-24-69 Guthrie	52,870	52,870
	1	85790 W motor	1.07	Fund. of Cost Engr. 1964	3,970	3,970
6. Slurry feed tank	1	Capacity 20.9 m^3, carbon steel	0.68	Chem. Engr. 3-24-69 Guthrie	5,570	5,570
Lining	1	6.35×10^{-3} m neoprene	----	----	4,930	4,930
7. Agitator, slurry feed tank	1	1492 W, neoprene coated	0.50 0.46	Chem. Engr. 3-24-69 Guthrie Fund. of Cost Engr. 1964	3,020	3,020
8. Pumps, slurry feed tank	2	7.6×10^{-4} m^3/s, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		1,990	3,980
9. Dust collecting system	1	0.47 m^3/s, inertial separator, cyclone, hoppers, fan, and drive	0.80	Chem. Engr. 3-24-69 Guthrie	1,460	1,460
10. Hoist	1	1800 kg electric	0.81	Popper, H.	10,890	10,890
11. Bag filter system	1	0.94 m^3/s, automatic fabric dust collectors, bag support, shaker system, isolation damper, motor, drive, dust hopper, fan and motor	0.68	Chem. Engr. 3-24-69	2,730	2,730
SUBTOTAL						97,230

TABLE D-2 (Continued)
AREA 3 - PARTICULATE SCRUBBING

Item	No.	Description	Size-Cost Scale Factor	Factor Source	Base Cost Each (1977)	Total Mid-1977 Cost
1. Tank particulate scrubber, effluent hold	2	Capacity 199.4 m^3, carbon steel	0.68	Chem. Engr. 3-24-69 Guthrie	31,970	63,940
Lining	2	6.35×10^{-3} m neoprene	----	----	23,540	27,080
2. Agitator, effluent hold tank	2	7460 W, neoprene coated	0.26 0.50	Fund. of Cost Engr. 1964 Chem. Engr. 3-24-69 Guthrie	5,770	11,450
3. Pumps, recycle slurry	3	.26 m^3/s, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		17,450	52,350
4. Venturi scrubber	2	58.7 m^3/s, carbon steel, neoprene lined	0.60	Universal Oil Products	121,040	242,080
5. Venturi sump	2	Carbon steel, neoprene lining	0.68	Chem. Engr. 3-24-69 Guthrie	44,000	88,000
6. Soot blowers	10		1.00	TVA	4,820	48,200
7. Bleed pump	3	1.6×10^{-3} m^3/s, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		2,000	6,000
SUBTOTAL						539,190

TABLE D-2 (Continued)

AREA 4 - SO_2 SCRUBBING

Item	No.	Description	Size-Cost Scale Factor	Factor Source	Base Cost Each (1977)	Total Mid-1977 Cost
1. Spray tower scrubber	2	Gas Flow 117.4 m^3/s, carbon steel, neoprene		Wester Precipitator Div., Joy Mfg. Co.[a]	145,260	290,520
2. Spray tower sump	2	Carbon steel, neoprene lined	0.68	Chem. Engr. 3-24-69 Guthrie	44,000	88,000
3. Tank, absorber effluent hold	2	Capacity 707.9 m^3, carbon steel, field erected	0.68	Chem. Engr. 3-24-69 Guthrie	45,820	91,640
Lining	2	6.35×10^{-3} m neoprene	----	----	39,220	78,440
4. Agitator, SO_2 absorber hold tank	2	29840 W, neoprene coated	0.50	Chem. Engr. 3-24-69 Guthrie	13,250	
5. Pumps, SO_2 absorber recycle	5	.46 m^3/s, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		27,000	27,000
6. Pumps, makeup water	2	1.2×10^{-3} m^3/s, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		1,500	3,000
7. Soot blowers	10		1.00	TVA	4,820	48,200
8. Demister	2	Carbon steel, neoprene lined	----	----	14,500	29,000
9. Pump, bleed	4	6.7×10^{-4} m^3/s, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		2,000	8,000
10. Tank, demister wash	2	Capacity 1.89 m^3, carbon steel, neoprene lined	0.68	Chem. Engr. 3-24-69 Guthrie	1,400	2,800
11. Pump, demister wash	4	1.3×10^{-3} m^3/s, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		1,500	6,000
SUBTOTAL						672,600

---

[a] Indicates source of spray tower cost

TABLE D-2 (Continued)

AREA 5 - REHEAT

Item Item	No.	Description	Size-Cost Scale Factor	Factor Source	Base Cost Each (1977)	Total Mid-1977 Cost
1. Steam reheater	2	2.0×10^6 W rating 73.4 m^2 surface area	0.80	Chem. Engr. 3-24-69	52,350	104,700
2. Soot	10		1.00	TVA	4,820	48,200
SUBTOTAL						152,900

AREA 6 - GAS HANDLING

Item	No.	Description	Size-Cost Scale Factor	Factor Source	Base Cost Each (1977)	Total Mid-1977 Cost
1. Fan	2	1.14×10^6 W drive	0.68	Chem. Engr. 3-24-69 Guthrie	51,430	102,860

AREA 7 - SOLIDS DISPOSAL

1. Clarifier	1	5.8×10^3 m^3/s	0.68	PEDCO (PE-146)	171,600	171,600
2. Pumps, pond feed	2	1.4×10^{-3} m^3/s, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		1,500	3,000
3. Pump, clarifier water recycle	2	4.4×10^{-3} m^3/s, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		3,500	7,000
4. Pumps, particulate pond water recycle	2	3.7×10^{-3} m^3/s, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		4,000	8,000
5. Pumps, SO2 pond water recycle	2	5.8×10^{-4} m^3/s, carbon steel, neoprene lined			1,100	2,200
SUBTOTAL						191,800

TABLE D-2 (Continued)

AREA 8 - UTILITIES

Note: There is no process equipment in this area.

AREA 9 - SERVICES

Item	No.	Description	Size-Cost Scale Factor	Factor Source	Base Cost Each (1977)	Total Mid-1977 Cost
1. Payloader			---	----	29,850	29,850
2. Plant vehicles			---	----	----	12,050
3. Maint. & instrument shop-equipment			---	----	31,780	31,780
4. Service building-equipment			---	----	42,130	42,130
5. Stores-equipment			---	----	12,760	12,760
SUBTOTAL						128,570

AREA 10 - PARTICLE RECIRCULATION

Item	No.	Description	Size-Cost Scale Factor	Factor Source	Base Cost Each (1977)	Total Mid-1977 Cost
1. Wet ball mill	1	3.2×10^{-4} m^3/s	.65	McGlamery	29,950	29,950
2. Pump, particle recirculation	2	3.2×10^{-4} m^3/s, molded polypropylene	Depends on gpm and head requirements resulting in changes of motor and impeller size		500	1,000
3. Tank, particle recirculation surge	1	Capacity 1.1 m^3, carbon steel, neoprene lined	.68	McGlamery	1,000	1,000
SUBTOTAL						31,950

2.0 COST FOR SLUDGE PONDS

The cost for the sludge ponds was not included in the previous section because they were not considered to be equipment items. The cost of the ponds, unlike the equipment items, includes installation, and the cost for pumps and piping to and from the pond (~1 mile) are included. The ponds are clay-lined, and they are sized for a 30-year operation at 7000 hr/yr operating time. A midwestern plant location is assumed. The total cost is expected to vary for each specific plant location due to differing land costs. Table D-3 contains the cost information for the sludge ponds.

TABLE D-3

COST DATA FOR SLUDGE PONDS

Item	Area @ 40 foot Depth (acres)	Source	Total Cost Mid-1977 Dollars
Prescrubber settling pond, Standard System	8.2	TVA	120,000
Absorber settling pond, Standard System	11.5	TVA	170,000
Prescrubber settling pond, Recycle System	7.1	TVA	104,000
Absorber settling pond, Recycle System	11.2	TVA	166,000